U.S. DEPARTMENT OF COMMERCE
National Technical Information Service
PB80-150873

Technology Assessment Report for Industrial Boiler Applications Flue Gas Desulfurization

Radian Corp, Durham, NC

Prepared for

Industrial Environmental Research Lab, Research Triangle Park, NC

Nov 79

United States
Environmental Protection
Agency

Industrial Environmental Research
Laboratory
Research Triangle Park NC 27711

EPA-600/7-79-178i
November 1979

PB80-150873

EPA

Technology Assessment Report for Industrial Boiler Applications:
Flue Gas Desulfurization

Interagency Energy/Environment R&D Program Report

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NATIONAL TECHNICAL
INFORMATION SERVICE
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SPRINGFIELD, VA. 22161

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TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO.
EPA-600/7-79-178i

2.

3. RECIPIENT'S ACCESSION NO.
PB80 150873

4. TITLE AND SUBTITLE
Technology Assessment Report for Industrial Boiler Applications: Flue Gas Desulfurization

5. REPORT DATE
November 1979

6. PERFORMING ORGANIZATION CODE

7. AUTHOR(S)
J. C. Dickerman and K. L. Johnson

8. PERFORMING ORGANIZATION REPORT NO.

9. PERFORMING ORGANIZATION NAME AND ADDRESS
Radian Corporation
P.O. Box 8650
Durham, North Carolina 27707

10. PROGRAM ELEMENT NO.
EHE624

11. CONTRACT/GRANT NO.
68-02-2608, Task 47

12. SPONSORING AGENCY NAME AND ADDRESS
EPA, Office of Research and Development
Industrial Environmental Research Laboratory
Research Triangle Park, NC 27711

13. TYPE OF REPORT AND PERIOD COVERED
Task Final: 6/78 - 10/79

14. SPONSORING AGENCY CODE
EPA/600/13

15. SUPPLEMENTARY NOTES IERL-RTP project officer is John E. Williams, Mail Drop 61, 919/541-2483.

16. ABSTRACT
The report gives results of an assessment of the applicability of flue gas desulfurization (FGD) technology to industrial boilers and is one of a series to aid in determining the technological basis for a New Source Performance Standard for Industrial Boilers. The development status and performance of alternative FGD control techniques were assessed and the cost, energy, and environmental impacts of the most promising were identified. The study concluded that there is no best FGD technology for application to industrial boilers: each alternative has advantages and disadvantages which could make it best for a specific application. Cost estimates of applying FGD processes indicated that the cost effectiveness varies significantly depending on the fuel fired, boiler size, and control level. However, boiler size is the most significant factor affecting cost effectiveness: the economy of scale causes control of large sources to be the most effective. The energy requirement of applying FGD processes varied from about 0.5% to 6% of boiler capacity, excluding stack gas reheat. The environmental impacts of each alternative were evaluated: each could be applied in an environmentally acceptable manner under existing regulations. The report does not consider combinations of technology to remove all pollutants, and these findings have not undergone detailed assessments for regulatory action.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution Flue Gases Desulfurization Assessments Boilers	Pollution Control Stationary Sources Industrial Boilers	13B 21B 07A, 07D 14B 13A

18. DISTRIBUTION STATEMENT Release to Public	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

Form 2220-1 (9-73)

EPA-600/7-79-178i

November 1979

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Flue Gas Desulfurization

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Contract No. 68-02-2608
Task No. 47
Program Element No. EHE624

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U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Research and Development
Washington, DC 20460

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PREFACE

The 1977 Amendments to the Clean Air Act required that emission standards be developed for fossil-fuel-fired steam generators. Accordingly, the U.S. Environmental Protection Agency (EPA) recently promulgated revisions to the 1971 new source performance standard (NSPS) for electric utility steam generating units. Further, EPA has undertaken a study of industrial boilers with the intent of proposing a NSPS for this category of sources. The study is being directed by EPA's Office of Air Quality Planning and Standards, and technical support is being provided by EPA's Office of Research and Development. As part of this support, the Industrial Environmental Research Laboratory at Research Triangle Park, N.C., prepared a series of technology assessment reports to aid in determining the technological basis for the NSPS for industrial boilers. This report is part of that series. The complete report series is listed below:

Title	Report No.
The Population and Characteristics of Industrial/ Commercial Boilers	EPA-600/7-79-178a
Technology Assessment Report for Industrial Boiler Applications: Oil Cleaning	EPA-600/7-79-178b
Technology Assessment Report for Industrial Boiler Applications: Coal Cleaning and Low Sulfur Coal	EPA-600/7-79-178c
Technology Assessment Report for Industrial Boiler Applications: Synthetic Fuels	EPA-600/7-79-178d
Technology Assessment Report for Industrial Boiler Applications: Fluidized-Bed Combustion	EPA-600/7-79-178e
Technology Assessment Report for Industrial Boiler Applications: NO_x Combustion Modification	EPA-600/7-79-178f
Technology Assessment Report for Industrial Boiler Applications: NO_x Flue Gas Treatment	EPA-600/7-79-178g
Technology Assessment Report for Industrial Boiler Applications: Particulate Collection	EPA-600/7-79-178h
Technology Assessment Report for Industrial Boiler Applications: Flue Gas Desulfurization	EPA-600/7-79-178i

These reports will be integrated along with other information in the document, "Industrial Boilers - Background Information for Proposed Standards," which will be issued by the Office of Air Quality Planning and Standards.

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ACKNOWLEDGMENT

The authors wish to acknowledge the key contributions made by the following people in the conduct of this evaluation. Overall guidance and coordination of this evaluation with other eveluations of industrial boiler control systems being conducted by EPA's Industrial Environmental and Research Laboratory (IERL) was performed by Mr. J. David Mobley. Mr. Charles Sedman provided overall project guidance for EPA's Office of Air Quality Planning & Standards (OAQPS). In addition, valuable input was received from the reviews of draft materials by several owners and operators of industrial boiler flue gas desulfurization systems, and their contribution is greatly appreciated.

SECTION 1
EXECUTIVE SUMMARY

1.1 INTRODUCTION

1.1.1 Background and Objective

The Clean Air Act Amendments of 1977 require the Environmental Protection Agency to coordinate and lead the development and implementation of regulations on air pollution. These include standards of performance for new and modified sources of pollution. Specifically mentioned in the Act are fossil fired steam generators. Accordingly, EPA has undertaken a study of industrial boilers with intent to establish emission control levels based upon the results of this and other studies.

This report presents the results of a study conducted to evaluate the applicability of various flue gas desulfurization (FGD) technologies for treating SO_2 emissions produced from small sized industrial boilers. Results of this evaluation will be used by the EPA to establish regulatory alternatives for small industrial boilers. Factors that were considered in evaluating the applicability of FGD technologies to industrial boilers included development status, capital and operating costs, energy requirements, environmental impacts, and performance and operating data.

1.1.2 Approach

In order to complete the objective of this study, a multiphased project approach was used. First, a comprehensive list of FGD processes was reviewed which included processes in commercial use, processes under development, and processes for which development efforts have ceased. Process status reports were prepared for eleven of these processes; those which are currently commercially used or are undergoing major demonstration efforts. Status reports for each process included detailed process descriptions, and discussions of design considerations and process performances.

The second phase involved selecting the candidate processes that appeared to be best suited to industrial boiler applications from the list of eleven processes for which the process status reports were prepared. These candidate technologies were compared using three emission control levels labelled "moderate, intermediate, and stringent" which correspond to SO_2 removal levels of 75, 85 and 90 percent. The control levels were chosen to form a basis of comparison of the control technologies considering performance, costs, energy, and non-air environmental impacts. A series of material and energy balance calculations were then prepared for the candidate processes to assess the environmental and energy impacts associated with process size, fuel sulfur content, and percent SO_2 removal. Finally, a series of capital and operating cost estimates were performed to assess the economic impact of the candidate FGD processes as functions of the above three variables.

From these comparisons, candidate "best technologies"for control of SO_2 were recommended for consideration in subsequent industrial boiler studies. These "best technology" recommendations do not consider combinations of technologies to remove more than one pollutant and have not undergone the detailed environmental, cost, and energy impact assessments necessary for regulatory action. Therefore, the levels of "moderate, intermediate, and stringent" and the recommendation of "best technology" for individual pollutants are not to be construed as indicative of the regulations that will be developed for industrial boilers. EPA will perform rigorous examination of several comprehensive regulatory options before any decisions are made regarding the standards for emissions from industrial boilers.

1.1.3 Conclusions

The major conclusion of this study is that there is no "best" FGD process for application to industrial boilers. Each of the five candidate processes has its own advantages and disadvantages which could serve to make it the "best" process for a specific application.

For the small sized applications, 8.8 MW_t (30×10^6 Btu/hr), the sodium throwaway process has the lowest annualized costs, is the lowest energy user, and produces a small aqueous waste stream which is generally treated in existing wastewater treating facilities. Currently, this process is the process of choice since 102 industrial boiler FGD systems (over 75 percent) are of the sodium throwaway type. However, there may be some areas where this process cannot be used because of regulations on the discharge of dissolved solids. In those cases either the limestone or double alkali process would probably become the process of choice.

The limestone and double alkali processes are considered together as their costs and environmental impacts are very similar. The double alkali process will have lower energy requirements than the lime/limestone process due to its more alkaline liquor and lower liquor circulation rate. The major disadvantage of these processes is the necessity of sludge disposal.

The spray drying/baghouse FGD system was considered only for low sulfur coal application due to data availability. Capital costs of this process were greater than the other throwaway processes due primarily to the fact that baghouse costs were included as part of the FGD system. Spray drying annualized costs, however, compared quite well with the other processes due primarily to the decreased disposal costs associated with handling a dry waste product. If baghouse costs were not included as part of this process, it would become the least expensive alternative. Future applications of this technology will probably be dependent upon the results of the five units that are currently under construction on both industrial and utility boilers.

The Wellman-Lord process will probably have limited applications for the process sizes evaluated in this study. This is due to its process costs, complexity in comparison to the other systems, and energy requirements.

However, for some applications with a strong by-product market, or with severe environmental regulations limiting the disposal of waste products from the other processes, the Wellman-Lord process may be attractive.

1.1.4 Report Organization

Section 2 of this report presents the process status reports for the eleven commercial and developing FGD processes that were evaluated for their potential to control SO_2 emissions from industrial boilers. Each of the processes are described and discussed with regard to their development status, design considerations, operating procedures, and system performance. Section 3 presents the methodology and results of a process screening that was used to select the five candidate processes for best control system for industrial boiler applications.

Section 4 presents the results of the cost analysis and discusses the effects of process size, fuel sulfur content, and SO_2 removal level on process costs. Sections 5 and 6 present process energy requirements and environmental impacts, respectively. Information presented in these sections illustrates the effects of process size, fuel sulfur content, and SO_2 removal levels on process energy requirements and environmental impacts. Section 7 presents the performance data that were obtained from industrial boiler FGD systems and Section 8 presents the impacts of partial scrubbing for the candidate FGD systems.

There are also two Appendices included with this report. Appendix A presents the results of the material and energy balance calculations for each of the processes and illustrates the variation in process stream flow rates with unit size, SO_2 removal level, and fuel sulfur content. Appendix B presents the results of the capital and operating cost estimates.

The remainder of this section presents a summary of the results and highlights of the information presented in the rest of the report.

1.2 EMISSION REDUCTION FOR COAL-FIRED INDUSTRIAL BOILERS

There are currently some 100 FGD processes that are in various stages of development including processes in early developmental stages and those for which development efforts have ceased. Of these processes, there are five that are in commercial use today in the United Statues. In addition, there are six that are currently at the demonstration stage. It is felt that these eleven systems will be used for the majority of near-term FGD applications to both utility and industrial boilers. Consequently, they are discussed in detail in Section 2 of this report. Table 1.2-1 presents a summary of the development status and industrial boiler applicability of these eleven processes.

1.2.1 Selection of Candidates for Best Control System

In order to select the candidate control systems, a set of evaluation or screening criteria were established to provide an objective and consistent means of comparing the processes and to insure that the same factors were considered for each process. The screening criteria were then applied to each process and the results were compared and used to select the processes that appeared to be best suited for near-term industrial boiler applications. The criteria used for this screening are listed in Table 1.2-2.

TABLE 1.2-2. FLUE GAS DESULFURIZATION SCREENING CRITERIA

1. Status of Development
   - Overall Process Development
   - Availability of Data
2. Performance
   - SO_2 Removal
   - Reliability
3. Applicability
   - Simplicity
   - Flexibility
   - Controllability
   - By-Product Marketability
4. Economic Considerations
   - Capital Investment Costs
   - Operating Costs
5. Energy Considerations
   - Liquid Pumping Requirements
   - System Pressure Drop
   - Regeneration Energy
   - Requirement for Reducing Gas
6. Environmental Considerations
   - Multipollutant Control
   - Secondary Pollutant Emissions

Process	Development status	No. of operational plants Industrial	 Utility	Applicability to industrial boilers
Lime/Limestone	Commercial industrial and utility applications.	2	28	Generally applicable. Possible limitations due to solids disposal land requirements.
Double Alkali	Commercial industrial applications - a 280 MW_e utility application is planned.	21	-	Generally applicable. Has demonstrated relatively reliable operations. Possible limitations due to solids disposal land requirements.
Wellman-Lord	Commercial applications for tail gas treating. A 115 MW_e utility demonstration test has been completed.	-	2	Generally applicable. Process costs and complexity will limit applications to small boilers. Has demonstrated good reliability.
Magnesium Oxide	Commercial utility applications. No planned industrial applications.	-	1	Process complexity will limit applications for industrial boilers. Long term reliability not demonstrated.
Sodium Scrubbing	Commercial industrial and utility applications.	119	3	Generally applicable. Possible limitations due to sorbent availability and cost, and water treatment.
Spray Drying	Pilot-scale. Commercial industrial utility systems are under construction.	-	-	SO_2 removal may be limited for lime based high sulfur coal applications. System is generally applicable except for land requirements for solids disposal. High reliability is claimed but undemonstrated.
Citrate/Phosphate	1 MW pilot-scale. A 64 MW_e industrial boiler application is planned.	-	-	Applicability to small boilers will be limited by overall complexity and the need for a reducing gas to produce H_2S.

TABLE 1.2-1. (Continued)

Process	Development status	No. of operational plants		Applicability to industrial boilers
		Industrial	Utility	
Bergbau-Forschung/ Foster Wheeler	20 MW demonstration in U.S. and a 45 MW demonstration in Germany.	-	-	Applicability will be limited by overall complexity and the requirement for extensive solids handling equipment.
Atomics International/ Aqueous Carbonate Process	1.25 MW nonintegrated pilot plant. A 100 MW utility demonstration is planned.	-	-	Applicability will be limited by overall complexity for small boiler applications. Use of unfamiliar technology in the reducing reactor may hinder process acceptability.
Shell/UOP	0.6 MW pilot plant in U.S. on coal-fired boiler. 40 MW in Japan on oil-fired boiler.	-	-	Applicability will be limited by overall complexity and the requirement for hydrogen for regeneration.
Chiyoda 121	Small-scale pilot plant. A 20 MW utility demonstration is planned.	-	-	Generally applicable. Possible limitations due to solids disposal land requirements in cases where by-product gypsum marketability is not feasible.

Results of the process screening were that five FGD processes were selected as candidate systems for application to industrial boilers. The processes selected were:

- Sodium Scrubbing
- Double Alkali
- Lime/Limestone
- Spray Drying
- Wellman-Lord

The first three processes are all currently used to control SO_2 emissions from industrial boilers throughout the United States. These processes are the ones of choice as evidenced by the fact that 118 of the 132 operating industrial boiler FGD systems are of these process types. The remaining operating industrial boiler FGD systems use ammonia process waste waters as a sorbent and are predominately found in sugar processing plants. Lime/limestone, double alkali, and sodium scrubbing processes also appear to be the processes of choice for future installations as evidenced by the fact that 36 out of 39 systems in the planning or construction stages are of these process types.

The spray drying process was selected as a candidate technology for industrial boiler applications due to its potential for widespread use as evidenced by the large amount of interest expressed in this rapidly developing process. Presently, there are no full-scale spray drying FGD systems in operation; however, orders have been placed for five commerical spray drying systems, two of which are industrial boiler applications. Data on spray drying is available only from pilot scale units. This data will be used in later sections to compare the energy, environmental, and economic impacts of this process versus the other candidate processes.

The Wellman-Lord process was selected to compare the impacts of a regenerable FGD process against the other candidate processes which are all throwaway systems. There are currently no regenerable systems in operation on small industrial boilers, however, a demonstration of the Citrate Process is scheduled for operation in the near future. The Wellman-Lord process was selected as a candidate process over the Citrate Process primarily because of its increased development status and availability of data.

1.2.2 Cost Analysis of Candidate Systems

Process costs are based on mid-1978 dollars and were evaluated as a function of process size, fuel sulfur content, and SO_2 removal for the five candidate FGD processes. The general approach used in developing the process costs consisted of four main steps. First, a series of material and energy balance calculations were performed for each process to define process stream flow rates and energy requirements as functions of unit size, SO_2 removal, and the amount of sulfur in the coal. The energy requirements for each process are presented and discussed in Section 5 and the material balance results are presented in Appendix A. Second, each of the FGD processes were divided into a number of process modules which represented separate processing areas. Third, equipment sizes were then developed for each process module based on the results of the material and energy balances. Finally, capital cost estimates were prepared by contacting process equipment vendors for price quotations in the size range of the standard industrial boilers for this study. Except for the spray drying process, particulate control equipment costs were not included in this study.

The wastewater treating processing area for the sodium throwaway process, consisting of oxidation and pH neutralization, was assumed to be associated with the boiler or plant in question. Therefore, wastewater treatment appears as an operating cost only. Similarly, solids disposal for the other nonregenerable processes was assumed to be contracted out for offsite disposal and it, too, appears only as an operating cost.

Direct capital investment costs for the FGD processes ranged from a low of \$187,000 for a sodium throwaway process*(8.8 MW_t ($30x10^6$ Btu/hr), 75 percent removal, 0.6 percent sulfur) to a high of \$2,573,000 for a Wellman-Lord process (58.6 MW_t ($200x10^6$ Btu/hr), 90 percent removal, 3.5 percent sulfur coal). When indirect capital expenses are added, the total capital investment costs for these two cases become \$394,000 and \$4,233,000 respectively. From a capital cost standpoint, the sodium throwaway process appears to be the least costly process, and the Wellman-Lord the most costly process for all of the cases considered.

With regard to annualized costs, the relative rankings of the FGD processes remain the same as with the capital costs for all cases considered. The sodium throwaway process again emerged as the least costly alternative. It should be noted however, that part of the low costs for this process are attributable to the relatively simple waste treating process consisting of oxidation plus pH neutralization. If a more elaborate water treating scheme were required, possibly to comply with zero discharge regulations, process costs would increase accordingly and could even make this process the most costly alternative. Figures 1.2-1 and 1.2-2 illustrate graphically the relative capital and annualized costs of the candidate processes applied to boilers burning a 3.5 percent sulfur coal and Figures 1.2-3 and 1.2-4 illustrate the same costs for 0.6 percent sulfur coal applications.

The cost effectiveness of the various FGD processes was also determined as part of this study. Cost effectiveness was defined as dollars per kilogram of removed SO_2 (\$/kg SO_2) and was calculated by dividing the annualized process costs by the kilograms of SO_2 removed in a year assuming a 60 percent load factor. Results of these calculations show that both coal sulfur content and process size significantly affect the cost effectiveness of an FGD process.

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*Throughout this report, boiler sizes will be expressed as watts of heat input. To avoid confusion with electrical output watts, heat input energy will be expressed as MW_t and electrical output energy will be expressed as MW_e.

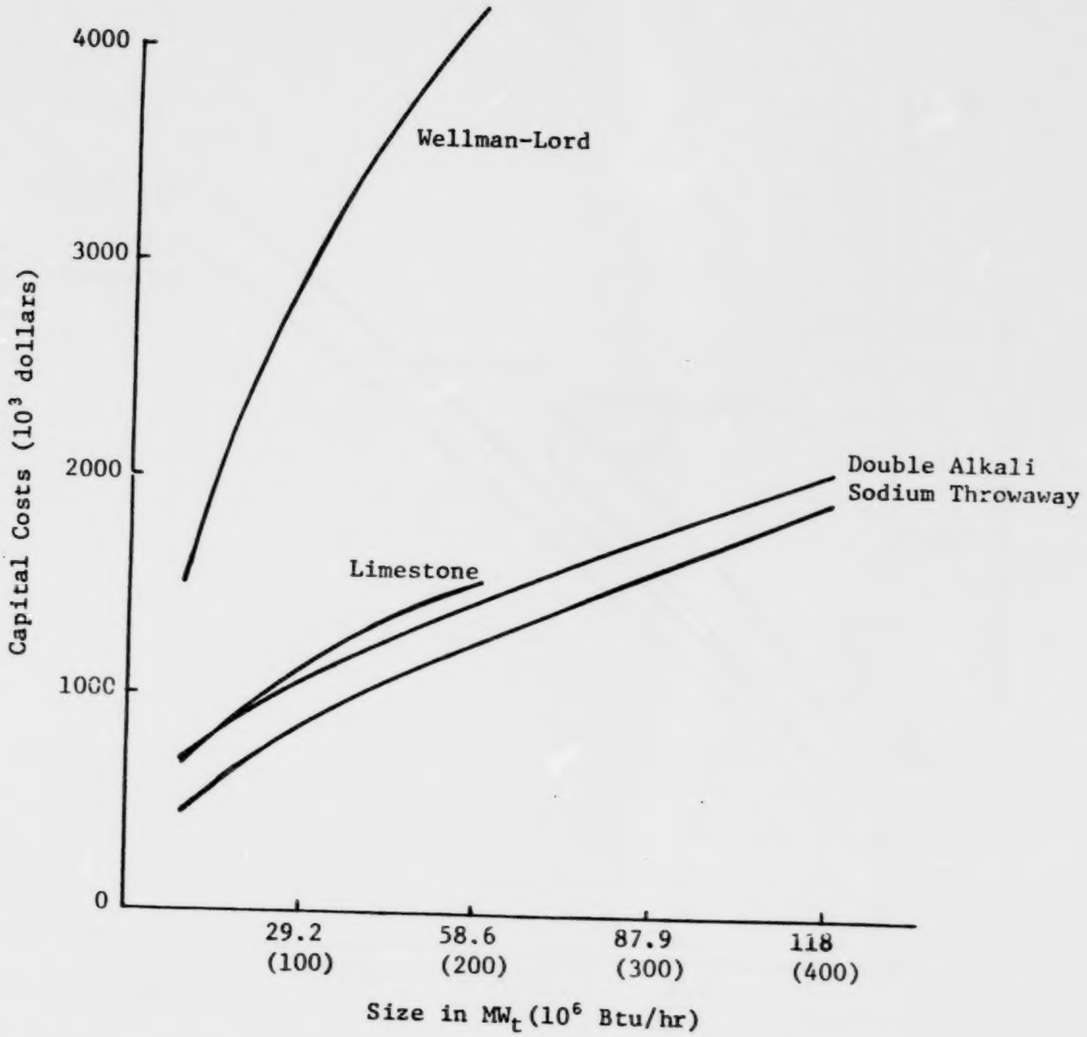


Figure 1.2-1. FGD capital costs *versus* unit size. (3.5% S coal, 90% removal)

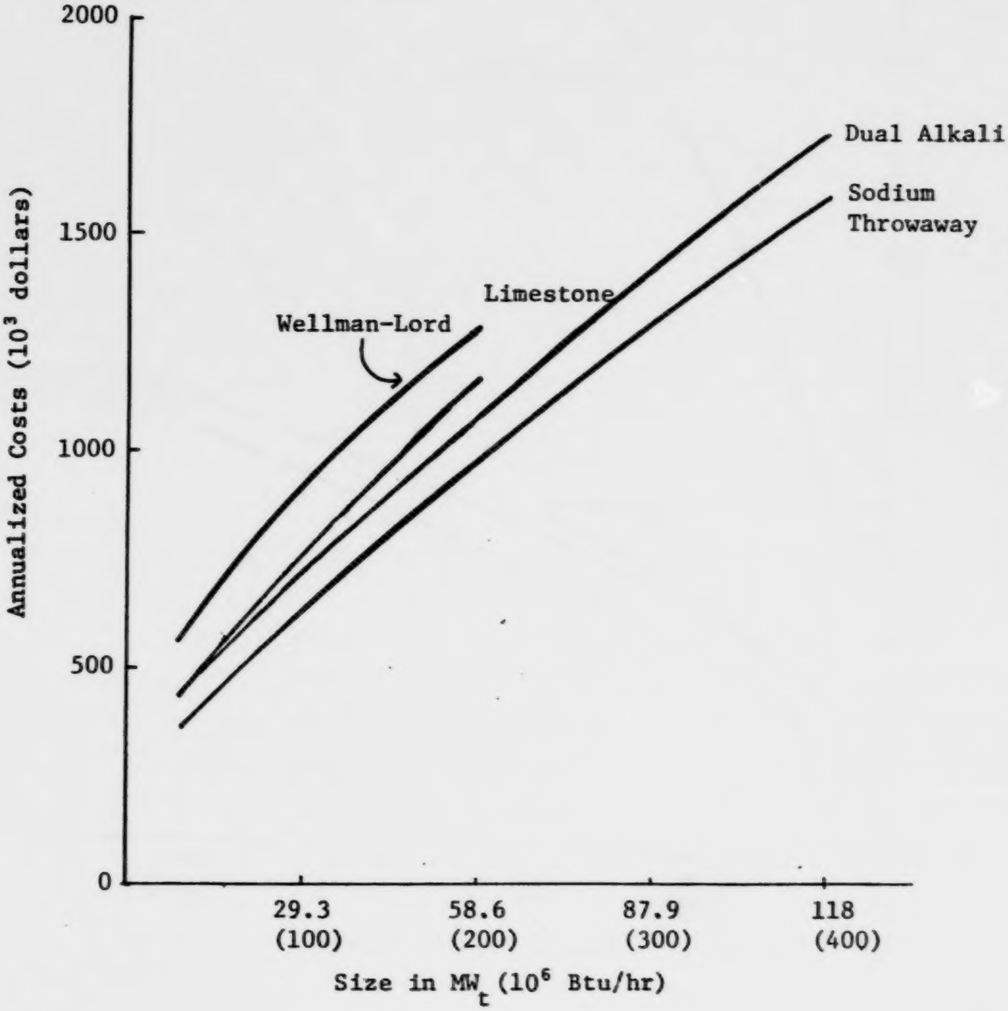


Figure 1.2-2. FGD annualized costs *versus* unit size. (3.5% S coal, 90% removal)

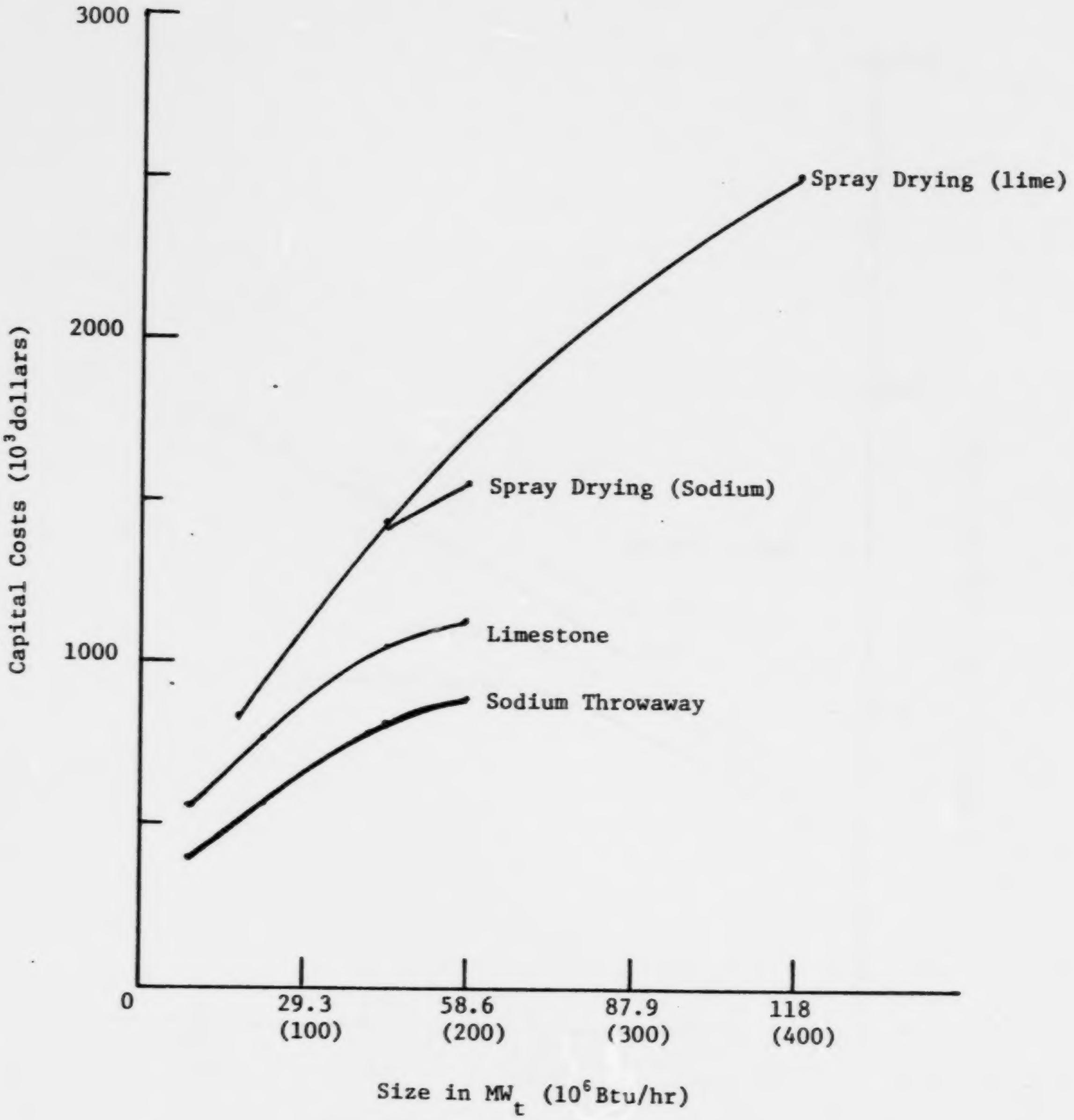


Figure 1.2-3. FGD capital costs *versus* unit size.
(0.6% S coal, 75% removal)

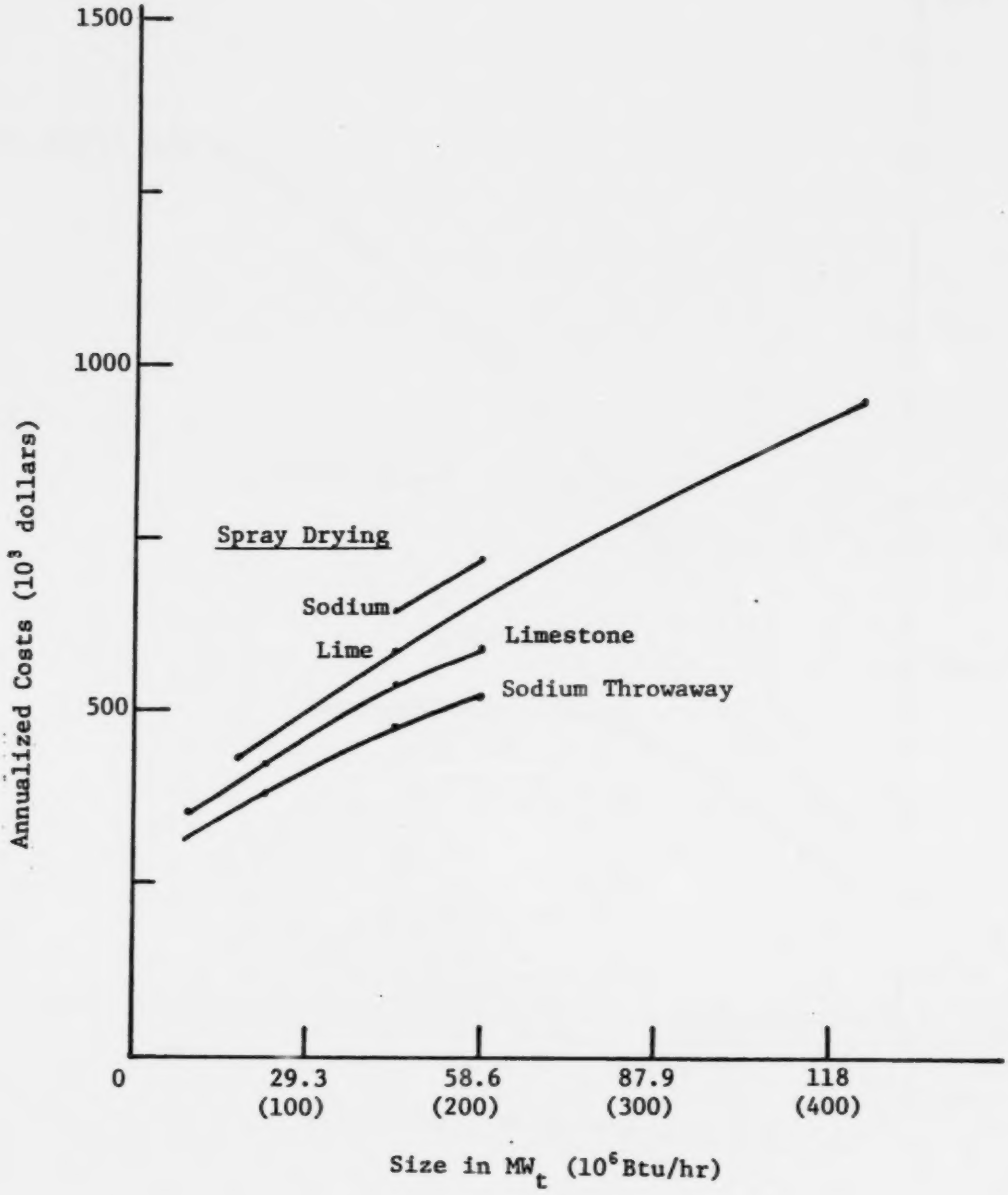


Figure 1.2-4. FGD annualized costs *versus* unit size. (0.6% S coal, 75% removal)

For a given size system, cost effectiveness increases with an increasing coal sulfur content. For a fixed coal sulfur content, cost effectiveness increases with increasing process size. Consequently, the most cost effective systems are those designed for the large sized boilers burning high sulfur coal, and the least cost effective systems are those designed for the small boilers burning low sulfur coal. Figure 1.2-5 illustrates these effects for the limestone processes. Curves developed for the other process showed similar effects.

1.2.3 Energy Impacts of Candidate Control Systems

Process energy requirements were evaluated as a function of process size, fuel sulfur content, and level of SO_2 control. Results of these calculations, shown in Table 1.2-3, indicate that the process energy penalties range from about 1/2 to 6 percent of the gross heat input to the boiler when no reheat is required, and from about 2 to 8 percent including stack gas reheat. Since the spray drying process exhausts its stock gas at 175^{o}F, no reheat energy penalties were charged for this process. The larger energy consumption for the Wellman-Lord process is due to the steam and methane requirements for the regeneration and SO_2 reduction portions of the process.

A summary of the relative percentage of the energy requirements of each process as compared to the overall energy requirement for the throwaway FGD processes applied to a boiler burning 3.5 percent sulfur coal is presented in Table 1.2-4. This table shows that fans are the largest energy consumer for each of the wet throwaway processes when stack gas reheat energy requirements are not included. However, when reheat energy requirements are included, they become the dominant energy consuming portion of these processes. The variations in energy requirements for these processes are due to different levels of sulfur in the coal, different levels of SO_2 control, and to a smaller extent, unit size.

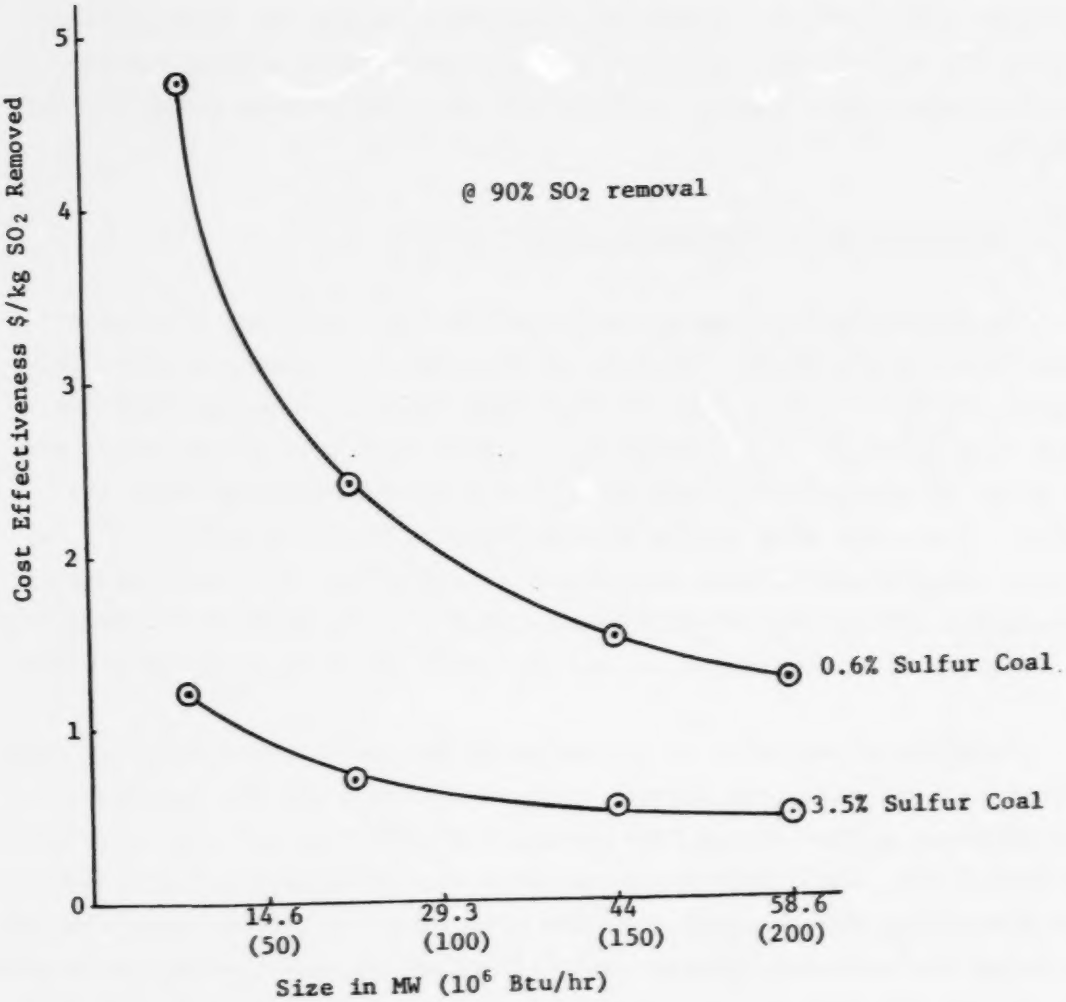


Figure 1.2-5. Limestone process cost effectiveness

TABLE 1.2-3. RANGE OF FGD SYSTEM ENERGY REQUIREMENTS*

SO_2 control method	Energy Requirement Not Including Reheat	Energy Requirement Including Reheat
Limestone	0.9-1.8	2.4-3.5
Sodium Scrubbing	0.5-0.8	2.0-2.6
Double Alkali	0.5-0.6	2.0-2.3
Spray Drying	0.5-0.8	0.5-0.8
Wellman-Lord	1.6-6.2	3.2-7.9

* Energy Requirements expressed as percent of net heat input to boiler.

TABLE 1.2-4. PERCENTAGE ENERGY CONSUMPTIONS FOR NONREGENERABLE PROCESSES ($58.6\ MW_t$ Boiler, 90% Removal, Eastern Coal)

Source of energy consumption	Limestone		Double Alkali		Sodium TA	
	kW_t	Percent of total	kw_t	Percent of total	kw_t	Percent of total
Raw materials handling and preparation	114.2	13	12.7	1	9.3	2
Liquid pumps	317.4	35	63.2	21	40.7	9
Fans	471.7	51	222.0	72	222.0	51
Disposal pumps	-	-	-	-	160.4	36
Utilities and services	8.6	1	9.3	3	9.3	2
TOTAL	911.9		307.2		441.7	
Reheat Steam	884.0		884.0		884.0	
	1795.9		1191.2		1325.7	

Energy requirements for the spray drying process are quite low, and compare well with the other nonregenerable processes when no reheat is included. Table 1.2-5 illustrates this for a low sulfur western coal case. If reheat is required, the spray drying process has a significant advantage since it exhausts its flue gas at 175°F and requires no reheat.

TABLE 1.2-5. COMPARISON OF SPRAY DRYING PROCESS ENERGY REQUIREMENTS (44 MW_t, 0.6% S coal, 75% Removal)

Source of energy consumption	Limestone		Sodium TA		Lime Spray Drying	
	kW_t	Percent of total	kW_t	Percent of total	kW_t	Percent of total
Raw Material Handling	15	3	1	-	1.5	1
Liquid Pumping	133	30	29	11	0.5	-
Fans	292	65	195	78	140	52
Disposal Pumps	-	-	21	8	-	-
Atomization	-	-	-	-	120	44
Utilities & Services	8	2	9	3	8	3
Total	448		254		270	
Reheat steam	775		775		-	
	1223		1029		270	

For the Wellman-Lord process, the relative amounts of energy consumed by the various process areas varies depending upon the sulfur content of the coal being burned as shown in Table 1.2-6. However, for both the eastern 3.5 percent sulfur coal case and the western 0.6 percent sulfur coal cases, the regeneration and sulfur production areas are the major energy users. It is doubtful that the energy requirements of the regeneration processing area can be reduced since double effect evaporators were assumed for these calculations which are some 45 percent more energy efficient than single effect evaporators. If reheat energy requirements are included, they become the major energy consumer for the low-sulfur western coal case and serve to increase the energy requirements of both cases by about 900 kW_t.

TABLE 1.2-6. PERCENTAGE ENERGY CONSUMPTION FOR WELLMAN-LORD PROCESS (58.6 MW_t 90% SO_2 Removal)

Source of energy consumption	Western coal (0.6%S)		Eastern coal (3.5%S)	
	kW_t	Percent of total	kW_t	Percent of total
Raw materials handling and preparation	1.9	<1	9.3	<1
Pumps	42.2	4	82.9	2
Fans	208.6	22	205.5	6
Process steam	469.0	49	2220.0	62
Methane	219.0	23	1048.0	29
Utilities and services	10.2	1	9.6	<1
Total	950.9		3575.3	
Reheat steam	900.0	48.6	884.0	20
	1850.9		4459.3	

1.2.4 Environmental Impacts of Candidate Control Systems

The air, liquid, and solid waste impacts of the candidate processes were considered as functions of size, SO_2 removal level, and fuel sulfur content. With regard to air pollution, each of the candidate FGD processes has the capability of both particulate and SO_2 removal, but only SO_2 removal was considered for this report. The impact of all the candidate systems as far as SO_2 emissions is the same since each of the processes can be designed to achieve the same level of SO_2 control. Figure 1.2-6 illustrates this point graphically.

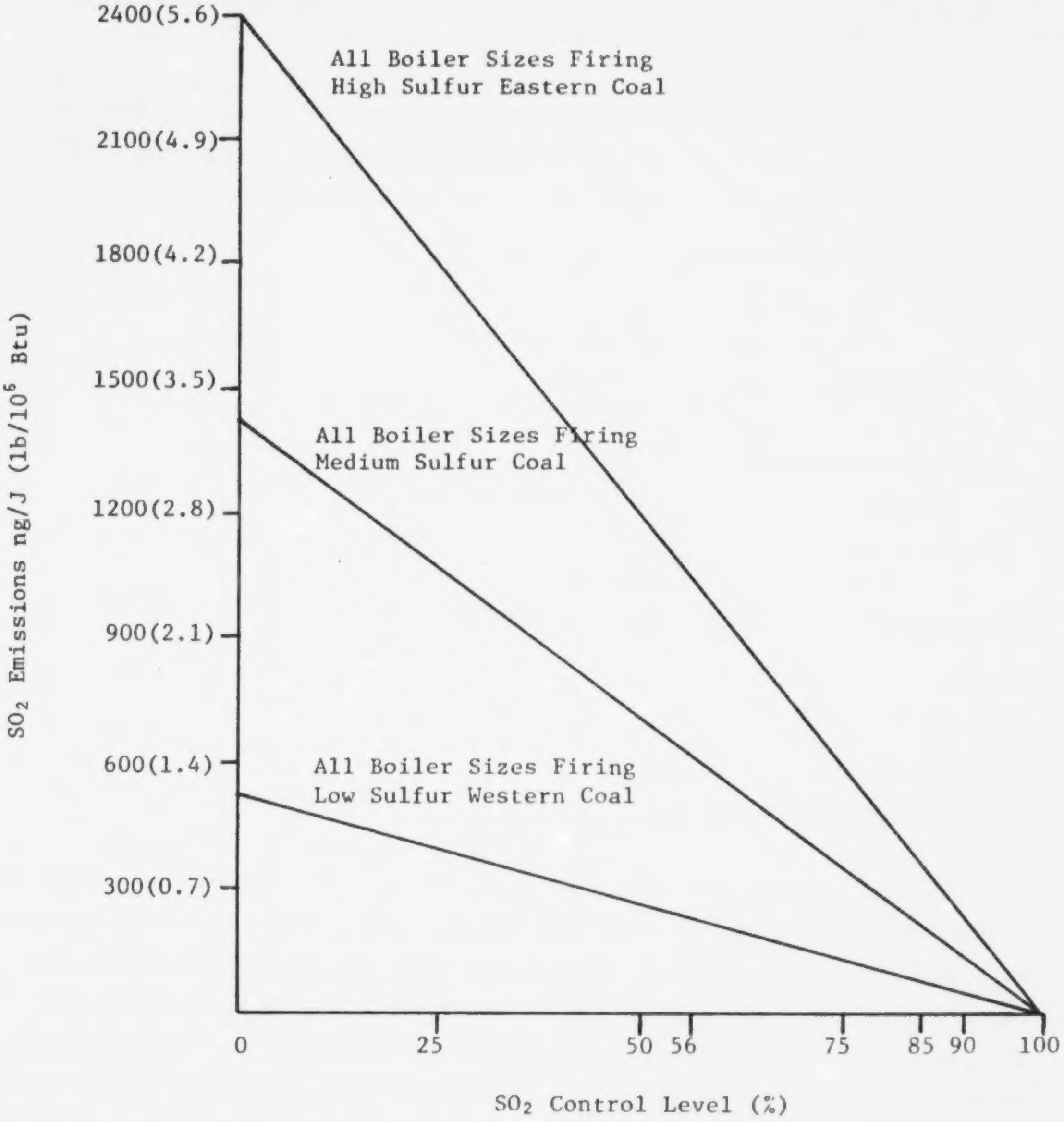


Figure 1.2-6. SO_2 emissions *versus* control level.

With regard to water pollution, only the sodium throwaway process should produce a significant environmental impact. Good design and operating practices for the limestone and double-alkali processes include dewatering the sludge and recycling the supernatant liquid. Dewatered sludges, however, contain up to 50 percent water which under proper design will be contained in the ultimate disposal site; be it a pond or landfill. Consequently, there should be essentially no water emissions from these systems except for times of process upsets.

The aqueous waste stream from the prescrubber of the Wellman-Lord process will be characterized by a low pH which results from the chlorides that are removed from the gas stream. However, except for the high chloride concentrations and low pH, the quality of the prescrubber discharge will be very similar to that of the boiler ash sluice water. Since this stream has been estimated to be approximately one percent of the ash sluicing requirements for a power plant, it can be used for ash sluicing where it will become diluted and neutralized with the other ash sluice water. Consequently, water emissions from the Wellman-Lord prescrubber stream should thus be limited to intermittent discharges from the ash pond.

The aqueous stream from a sodium throwaway system will contain about five percent dissolved solids. In these systems, the absorbed SO_2 reacts to form NA_2SO_3 and Na_2SO_4 which are removed from the system as dissolved solids in an aqueous waste. Consequently, the amount of aqueous emissions is directly related to both the SO_2 control level and the coal's sulfur concentration. Discharge rates and average stream compositions for the cases considered in this study are given in Table 1.2-7.

Common water treating practice for sodium throwaway systems in use today is to discharge their wastes to an evaporation pond or to an existing centralized water treating plant. Of the 102 sodium scrubbing systems in use today, about 80 use evaporation ponds and 10 use centralized water treating for disposal of their FGD wastes. The remaining systems use varied approaches ranging from discharge to city sewers to deep mine injection.

TABLE 1.2-7. AQUEOUS EMISSIONS FROM SODIUM THROWAWAY PROCESSES

Boiler size and type	Control level	3.5% S eastern coal ℓ/sec	3.5% S eastern coal (gpm)	2.3% S coal ℓ/sec	2.3% S coal (gpm)	0.6% S western coal ℓ/sec	0.6% S western coal (gpm)
8.8 MW_t(30 X 10^6 Btu/hr) Underfeed stoker	90	0.85	13.4	-	-	0.17	2.7
	85	0.79	12.5	-	-	0.16	2.6
	75	0.71	11.2	-	-	0.15	2.4
	56	0.52	8.4	-	-	-	-
22 MW_t(75 X 10^6 Btu/hr) Chaingrate Stoker	90	2.18	34.6	1.28	20.3	0.46	7.3
	75	1.77	28.1	-	-	0.37	5.9
	56	1.26	19.9	-	-	-	-
44 MW_t(150 X 10^6 Btu/hr) Spreader Stoker	90	4.38	69.4	-	-	0.92	14.6
	75	3.53	55.9	-	-	0.75	11.8
	56	2.45	38.8	-	-	-	-
58.6 Mw_t(200 X 10^6 Btu/hr) Pulverized coal	90	5.80	92.0	-	-	1.23	19.5
	85	5.26	83.4	-	-	1.11	17.6
	75	4.71	74.7	-	-	0.99	15.7
	56	3.34	52.9	-	-	-	-
118 MW_t(400 X 10^6 Btu/hr) Pulverized coal	90	11.7	185	6.82	108	2.46	38.9
Avg. Dissolved Solid Compositions		Na_2SO_3= 77 percent Na_2SO_4= 9 percent NA_2CO_3= 14 percent		73 percent 13 percent 14 percent		69 percent 17 percent 14 percent	
Avg. TDS Concentration (wt. %)		5		5		5	

*Based on material balance calculations provided in Appendix A.

For purposes of this evaluation, onsite treatment of sodium system aqueous wastes using a basic water treating scheme of sulfite oxidation and pH neutralization was selected as the treatment method for evaluation. Although evaporation ponds are currently used in the majority of sodium system applications, their use is limited to certain geographic areas of the country where the annual evaporation rate exceeds the annual rainfall. The water treatment system selected for this evaluation will result in a sodium sulfate stream which must be disposed.

The major solid waste impacts from the five candidate processes result from the sludges produced in the limestone and double-alkali processes and the dry solid produced in the spray drying process. This assumes, of course, that a market exists for the sulfur and H_2SO_4 produced in the Wellman-Lord process. A solid purge stream of Na_2SO_4 is also produced in the Wellman-Lord process, but the stream is relatively small and should not constitute a major solid waste impact, especially for the size applications under consideration in this evaluation.

Both the limestone and double alkali sludges are composed primarily of calcium sulfite and sulfate salts. Significant amounts of fly ash may also be present, depending on the method of upstream particulate control in use. For this study, upstream particulate removal was assumed and sludge production rates are given on an ash free basis. The sludges are relatively inert and can be disposed of in an environmentally acceptable manner. The disposal methods currently in use are lined and unlined ponding and landfilling of treated and untreated materials.

The dry solid waste product from the spray drying process will consist primarily of calcium or sodium salts, depending upon the type of alkali used as the SO_2 sorbent. Significant amounts of fly ash will also be present since the solids collection device associated with the spray drier, probably a baghouse, will remove the particulates generated from the coal combustion process along with the spray drying solid wastes. Upstream particulate removal is not practical for this process since the spray dryer's performance is not adversely affected by the presence of fly ash and dual particulate removal units would be unattractive from both an energy and economic view.

As with the sodium throwaway system, all of the SO_2 absorbed from the flue gas by a limestone system must leave the process in a waste stream, in this case as a waste sludge. Consequently, the amount of sludge produced by a limestone system is proportional to the sulfur content of the coal and the SO_2 removal level. Table 1.2-8 presents the results of the limestone process material balance calculations and shows the variation in sludge production with coal sulfur content and SO_2 removal. A 50 percent solids sludge was assumed for these calculations.

The volume of sludge produced is also important as the sludge volume will determine the size of the holding pond or landfill area. Table 1.2-9 presents the results of calculations to estimate the sludge volumes produced by a limestone process for the standard sized boilers. The difference in sludge densities for the two coals is due to the higher oxidation for the western coal cases. Results are presented in units of cc/sec, lb/hr, and acre-feet/30 years. The last category, acre-feet/30 years gives an indication of the total volume of sludge to be handled over the life of the plant assuming a 30 year life and an onstream factor of 60 percent. Figure 1.2-7 illustrates the results of these calculations graphically and shows the variation in sludge production with coal sulfur content, boiler size, and level of removal. As one would expect, sludge production increases with all of these factors.

The quantity of sludge produced from the double alkali process will also vary with sulfur removal and fuel sulfur content. Table 1.2-10 presents the solid waste impacts for the double-alkali process. The major difference between the amount of sludge produced from the double-alkali and limestone systems is that the limestone system stoichiometry is based upon 1.2 moles sorbent per mole SO_2 removed whereas the double-alkali system stoichiometry is based upon 1.0 moles sorbent per mole SO_2 removed.

TABLE 1.2-8. SOLID WASTE PRODUCTION FOR THE LIMESTONE FGD PROCESS
(Ash-Free Basis, 50 percent solids)

Boiler size and type	Percent removal	3.5% S eastern coal				0.6% S western coal			
		g/s	(lb/hr)	ℓ/min	(gal/min)	g/s	(lb/hr)	ℓ/min	(gal/min)
8.8 MW$_t$(30x10^6 Btu/hr)	90	89.8	(712)	2.7	(0.7)	20.9	(166)	0.61	(0.16)
Underfeed Stoker	85	84.7	(672)	2.7	(0.7)	19.4	(154)	0.57	(0.15)
	75	74.9	(594)	2.3	(0.6)	17.4	(138)	0.53	(0.14)
22 MW$_t$(75x10^6 Btu/hr)	90	226.5	(1796)	6.8	(1.8)	52.2	(414)	1.5	(0.4)
Chaingrate Stoker	75	190.2	(1508)	5.7	(1.5)	43.6	(346)	1.1	(0.3)
44 MW$_t$(150x10^6 Btu/hr)	90	457.0	(3624)	13.6	(3.6)	104.7	(830)	3.0	(0.8)
Spreader Stoker	75	380.6	(3018)	11.4	(3.0)	87.3	(692)	2.7	(0.7)
58.6 MW$_t$(200x10^{6}Btu/hr)	90	606.8	(4812)	18.2	(4.8)	139.5	(1106)	4.2	(1.1)
Pulverized Coal	85	576.1	(4568)	17.4	(4.6)	131.9	(1046)	3.8	(1.0)
	75	507.0	(4020)	15.2	(4.0)	116.0	(920)	3.4	(0.9)

TABLE 1.2-9. SOLID WASTE VOLUMES FOR THE LIMESTONE FGD PROCESS
(Ash-Free Basis, 50 percent solids)

Boiler size and type	Percent removal	3.5% S eastern coal* Sludge volume cc/sec	3.5% S eastern coal* Sludge volume (ft^3/hr)	3.5% S eastern coal* Sludge volume (acre-ft/ 30 yrs)	0.6% S western coal† Sludge volume cc/sec	0.6% S western coal† Sludge volume (ft^3/hr)	0.6% S western coal† Sludge volume (acre-ft/ 30 yrs)
8.8 MW_t(30x10^6 Btu/hr)	90	69.1	8.8	31.8	13.8	1.8	6.6
Underfeed Stoker	85	65.2	8.3	30.0	12.8	1.6	5.8
	75	57.6	7.3	26.4	11.5	1.5	5.4
22 MW_t(75x10^6 Btu/hr)	90	174.2	22.1	80.0	34.6	4.4	16.0
Chaingrate Stoker	75	146.3	18.6	67.4	28.9	3.7	13.4
44 MW_t(150x10^6 Btu/hr)	90	351.5	44.6	161.4	69.3	8.8	32.0
Spreader Stoker	75	292.8	37.2	134.6	57.8	7.4	26.8
58.6 MW_t(200x10^6 Btu/hr)	90	466.8	59.3	214.6	92.4	11.8	42.8
Pulverized coal	85	443.2	56.3	203.8	87.4	11.1	40.2
	75	390.0	49.5	179.2	76.8	9.8	35.4

*Eastern sludge bulk density = 1.30 g/cc (81.2 lb/ft^3) }

†Western sludge bulk density = 1.51 g/cc (94.0 lb/ft^3) } These are average values from Table 6.2-12.

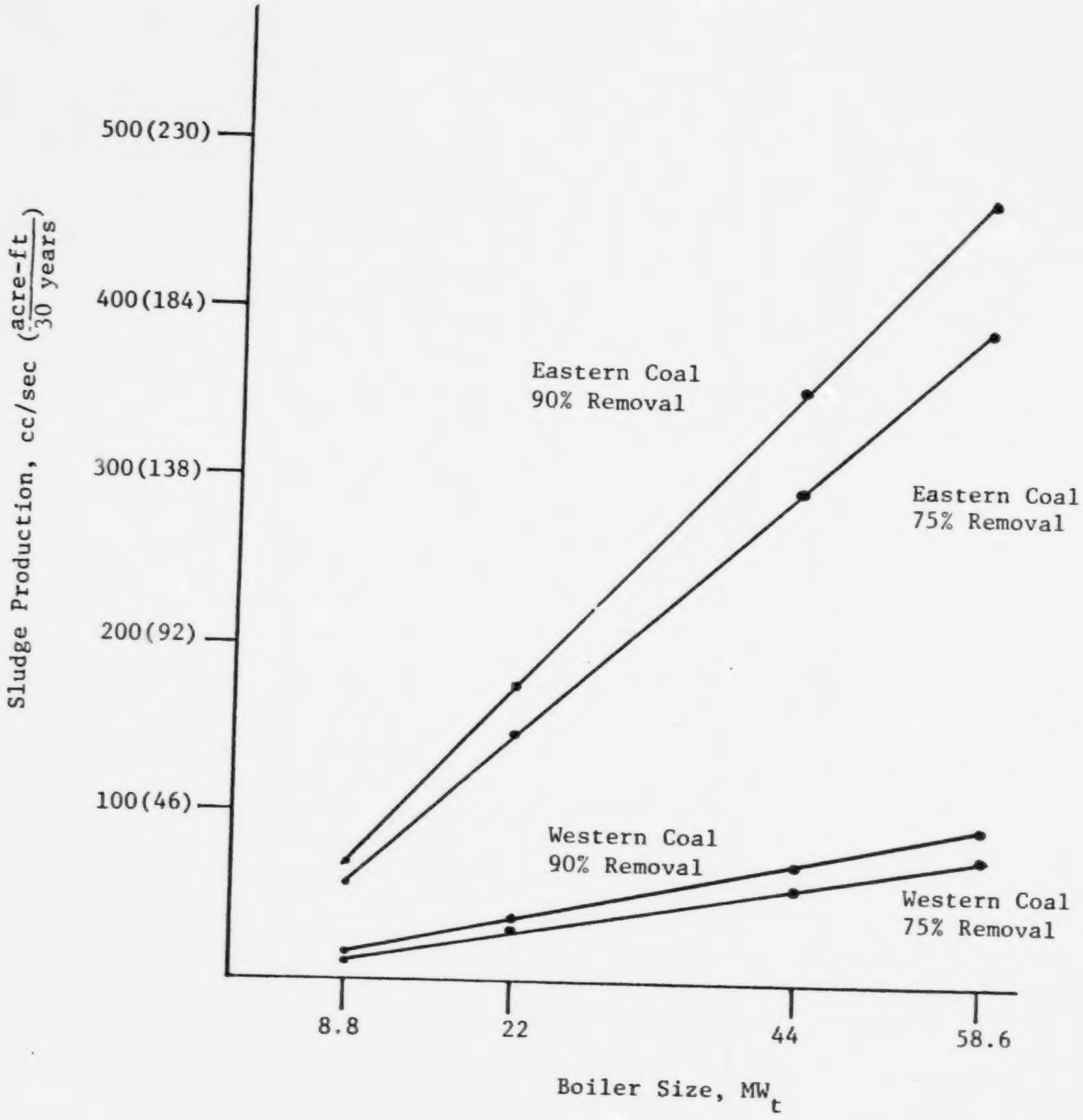


Figure 1.2-7. Sludge production rates for the limestone FGD process (50 percent solids in sludge)

TABLE 1.2-10. SOLID WASTE IMPACT FOR THE DOUBLE ALKALI PROCESS (Ash-free Basis, 50 percent solids)

Boiler Size and type	Percent removal	3.5% S eastern coal				0.6% S western coal				2.3% S coal			
		g/s	(lb/hr)	ℓ/min	(gal/min)	g/s	(lb/hr)	ℓ/min	(gal/min)	g/s	(lb/hr)	ℓ/min	(gal/min)
8.8 MW_t (30 X 10^6 Btu/hr) Underfeed Stoker	90	78.7	(624)	2.3	(0.6)	15.4	(122)	0.4	(0.1)	-	-	-	-
22 MW_t (75 X 10^6 Btu/hr) Chaingrate Stoker	90	200.0	(1584)	5.7	(1.5)	38.1	(302)	1.1	(0.3)	112.0	(888)	3.0	(0.8)
58.6 MW_t (200 X 10^6 Btu/hr) Pulverized Coal	90	506.0	(4012)	14.4	(3.8)	101.6	(806)	2.7	(0.7)	-	-	-	-
118 MW_t (400 X 10^6 Btu/hr) Pulverized Coal	90	1065.6	(8450)	30.7	(8.1)	203.5	(1614)	5.3	(1.4)	596.8	(4732)	17.1	(4.5)

All of the SO_2 absorbed in a spray dryer must also exit the process as a waste stream, in this case as a solid salt. Table 1.2-11 presents the quantity of solid wastes produced by the spray drying system as a function of SO_2 removal, coal type, and process size. Solid waste quantities from this process are, however, a combination of spray dryer solids and fly ash generated from coal combustion. It is interesting to note that for the cases considered in this evaluation, the majority of solid wastes from this process resulted from fly ash and not from the removal of SO_2. Table 1.2-12 presents a breakdown of the origin of solid waste material for each of the spray drier cases.

Results of pilot plant testing reported by Basin Electric were that the spray drying product produced from the coals they tested handled as well as fly ash and would not require special handling equipment other than the conventional dry handling equipment used for fly ash. Disposal methods planned for the systems they have under construction are disposal in depleted mines and landfill after mixing with conventional scrubber sludge. The two spray drying systems under construction for industrial boiler applications both plan to truck the waste solids to an offsite landfill area.

TABLE 1.2-11. SOLID WASTES FROM SPRAY DRYING
(Total Fly Ash + Alkali Salts)

MW_t	Boiler size (10^6 Btu/hr)	% S Coal	SO_2 Removal	Sorbent	Solids		Volume[a]		
					g/sec	(lb/hr)	cc/sec	(ft^3/hr)	acre-ft/15 years
44	(150)	0.6	90	Sodium	132	(1047)	111	(14.1)	25.5
44	(150)	0.6	75	Sodium	114	(904)	96	(12.2)	22.1
44	(150)	0.6	50	Sodium	99	(785)	83	(10.6)	19.2
58.6	(200)	0.6	75	Sodium	174	(1378)	146	(18.5)	33.5
17.6	(60)	0.6	75	Sodium	29	(227)	24	(3.1)	5.6
44	(150)	0.6	90	Lime	134	(1066)	113	(14.3)	25.9
44	(150)	0.6	75	Lime	114	(905)	96	(12.2)	22.0
44	(150)	0.6	50	Lime	96	(764)	81	(10.3)	18.6
118	(400)	0.6	70	Lime	344	(2725)	289	(36.7)	66.5
118	(400)	2.3	70	Lime	729	(5782)	613	(77.9)	141.0
22	(75)	2.3	70	Lime	85	(675)	71	(9.1)	16.5

a - Based on density = 1.19 g/cc (74.2 lb/ft^3) Reference 14

TABLE 1.2-12. SOLID WASTE BY ORIGIN

MW_t	(10^6 Btu/hr)	% S Coal	SO_2 Removal	Type sorbent	Amount of solid waste					
					Fly ash		Percent of total	Desulfurization products		Percent of total
					g/s	(lb/hr)		g/s	(lb/hr)	
44	(150)	0.6	90	Sodium	69	(546)	52	63	(501)	48
44	(150)	0.6	75	Sodium	69	(546)	60	43	(341)	40
44	(150)	0.6	50	Sodium	69	(546)	70	30	(239)	30
58.6	(200)	0.6	75	Sodium	113	(896)	65	61	(482)	35
17.6	(60)	0.6	75	Sodium	11	(84)	37	18	(143)	
44	(150)	0.6	90	Lime	69	(546)	51	66	(520)	49
44	(150)	0.6	75	Lime	69	(546)	60	45	(359)	40
44	(150)	0.6	50	Lime	69	(546)	71	27	(218)	29
118	(400)	0.6	70	Lime	226	(1791)	66	118	(934)	34
118	(400)	2.3	70	Lime	401	(3182)	55	328	(2600)	45
22	(75)	2.3	70	Lime	24	(187)	27	61	(488)	

SECTION 2

EMISSION CONTROL TECHNIQUES

2.1 PRINCIPLES OF CONTROL

There are several methods that may be used to reduce SO_2 emissions from industrial boilers to comply with current emission standards. SO_2 reduction methods that are applicable for a given site will be dependent upon the required SO_2 removal or emission levels. Potential SO_2 removal methods are: 1) use of low sulfur coal, 2) physical or chemical coal cleaning, 3) use of synthetic fuels from gasification or liquefaction plants, 4) fluid bed combustion, and 5) flue gas desulfurization. This report discusses flue gas desulfurization (FGD) as a method of controlling SO_2 emissions.

The sulfur content of the fuel fired in a boiler dictates the amount of SO_2 emissions produced at a given installation. Fuels of concern for this study are coal and residual oil. In general, light fuel oils and natural gas are treated to remove sulfur prior to combustion. Table 2.1-1 lists the proximate analyses for the standard fuels to be considered in this evaluation. This table also presents the relative amount of uncontrolled SO_2 emissions that would result from each assuming that 95 percent of the fuel's sulfur content is converted to SO_2.

There are currently some 100 FGD processes that are in various stages of development including processes in early developmental stages and those for which development efforts have ceased.[1] Of these processes, there are five that are in commercial use today in the United States. In addition, there are six that are currently at an advanced demonstration stage. Table 2.1-2 presents a summary of these eleven processes with regard to their development status and applicability to industrial boilers. It is felt that these

TABLE 2.1-1. SO_2 EMISSIONS FOR VARIOUS FUEL TYPES

	Fuel type				
	High sulfur eastern	Low sulfur eastern	Subbituminous western	Residual oil	Distillate oil
Composition[a]					
HHV kJ/kg	27,447	32,099	22,323	41,714[b]	38,706[b]
(Btu/lb)	(11,800)	(13,800)	(9,600)	(149,800)	(139,000)
Percent S	3.5	0.9	0.6	3.0	0.5
Percent ash	10.6	6.9	5.4	0.1	-
Uncontrolled SO_2 Emissions ($lb/10^6$ Btu)	5.61	1.24	1.18	3.04	0.48
SO_2 emissions relative to high sulfur eastern coal	1	0.22	0.21	0.54	0.08

[a] Composition based on data provided by PEDCo as a basis for this study.

[b] HHV expressed as kJ/kg (Btu/gal)

TABLE 2.1-2. FGD SYSTEM SUMMARY

Process	Development status	No. of operational plants		Applicability to industrial boilers
		Industrial	Utility	
Lime/Limestone	Commercial industrial and utility applications.	2	28	Generally applicable. Possible limitations due to solids disposal land requirements.
Double Alkali	Commercial industrial applications - a 280 MW_e utility application is planned.	21	-	Generally applicable. Has demonstrated relatively reliable operations. Possible limitations due to solids disposal land requirements.
Wellman-Lord	Commercial applications for tail gas treating. A 115 MW_e utility demonstration test has been completed.	-	2	Generally applicable. Process costs and complexity will limit applications to small boilers. Has demonstrated good reliability.
Magnesium Oxide	Commercial utility applications. No planned industrial applications.	-	1	Process complexity will limit applications for industrial boilers. Long term reliability not demonstrated.
Sodium Scrubbing	Commercial industrial and utility applications.	119	3	Generally applicable. Possible limitations due to sorbent availability and cost, and water treatment.
Spray Drying	Pilot-scale. Commercial industrial utility systems are under construction.	-	-	SO_2 removal may be limited for lime based high sulfur coal applications. System is generally applicable except for land requirements for solids disposal. High reliability is claimed but undemonstrated.
Citrate/Phosphate	1 MW pilot-scale. A 64 MW_e industrial boiler application is planned.	-	-	Applicability to small boilers will be limited by overall complexity and the need for a reducing gas to produce H_2S.

TABLE 2.1-2. (Continued)

Process	Development status	No. of operational plants		Applicability to industrial boilers
		Industrial	Utility	
Bergbau-Forschung/ Foster Wheeler	20 MW demonstration in U.S. and a 45 MW demonstration in Germany.	-	-	Applicability will be limited by overall complexity and the requirement for extensive solids handling equipment.
Atomics International/ Aqueous Carbonate Process	1.25 MW nonintegrated pilot plant. A 100 MW utility demonstration is planned.	-	-	Applicability will be limited by overall complexity for small boiler applications. Use of unfamiliar technology in the reducing reactor may hinder process acceptability.
Shell/UOP	0.6 MW pilot plant in U.S. on coal-fired boiler. 40 MW in Japan on oil-fired boiler.	-	-	Applicability will be limited by overall complexity and the requirement for hydrogen for regeneration.
Chiyoda 121	Small-scale pilot plant. A 20 MW utility demonstration is planned.	-	-	Generally applicable. Possible limitations due to solids disposal land requirements in cases where by-product gypsum marketability is not feasible.

eleven systems will be used for the majority of near-term FGD applications to both utility and industrial boilers. Consequently, they are evaluated in the following sections of this chapter.

Much of the data base for FGD systems, as presented in this chapter, has been collected for FGD systems applied to utility boilers as opposed to industry boilers. In general, utility operating data for FGD systems can be used to represent the operations of industrial boiler systems since the actual design of the boilers are very similar. However, there are some operating differences between utility and industrial boilers which may affect the operation of an FGD system when applied to industrial boilers. These will be discussed where appropriate in the following sections. Most notable of these are: 1) stoker-fired industrial boilers generally operate with higher quantities of excess air than pulverized coal boilers, 2) industrial boilers may have larger boiler load swings as they are operated to follow process operation, and 3) utility boiler operators are generally higher paid and better trained than industrial boiler operators. Another important difference is that when industrial boilers are down, replacement power or steam cannot generally be drawn from a grid as can be done for utility boilers. Consequently, a less complex, more reliable, and highly automated FGD system may be more suitable for industrial boiler applications.[4,5]

The higher amount of excess air used in stoker-fired boilers (may vary from 30-50 percent) will result in a higher oxygen content in the flue gas. Pulverized coal boilers for both utility and industrial applications are generally designed for 20-25 percent excess air. In some FGD systems, the higher oxygen concentration can adversely affect the rate of oxidation of the SO_2 scrubbing liquor. The dual alkali and Wellman-Lord systems are perhaps the most affected by this phenomenon. When the oxidation rate is high, significant amounts of sodium sulfate are formed in these systems which are difficult to regenerate and in many cases must be removed as a waste stream.

Wide load swings in industrial boilers are due to two main factors: 1) the need to follow process operations, and 2) load swings that occur due to the small boiler sizes. Boilers that provide steam for batch process operations may change load from near capacity to reserve standby in a matter of minutes. Apparently, this practice is dependent upon the industry and the site specific operations and no generalizations can be drawn. The small size of the industrial boilers also impacts their load swings since small boilers have less repetition of individual equipment. For example, in a small boiler with only two burners, if one of the burners goes out, a rapid loss of up to half the load can result. However, in a large boiler, loss of a single burner will not significantly affect the total load. The important point is that FGD systems applied to industrial boilers should be able to operate under rapidly changing load conditions.

2.2 CONTROLS FOR COAL-FIRED BOILERS

The eleven FGD systems summarized in Table 2.1-2 have been evaluated for possible application to control SO_2 emissions from small industrial boilers. These systems were evaluated as it was felt that they will be used for the majority of near-term FGD applications due to their relatively advanced state of development. The systems described in this chapter are listed below according to their current status of development.

1) Commercially Applied Processes

- Lime/Limestone
- Double Alkali
- Wellman-Lord
- Magnesium Oxide
- Sodium Scrubbing

2) Developing Processes

- Spray Drying
- Citrate/Phosphate
- Bergbau Forschung/Foster Wheeler

- Atomics International - Aqueous Carbonate
- Shell/UOP
- Chiyoda 121

The following discussions present a description of each of the above systems and evaluate them with regard to their development status, industrial boiler applicability, design and operating considerations, and system performance.

2.2.1 Lime/Limestone Wet Scrubbing

2.2.1.1 System Description--

A. System--The lime/limestone FGD process uses a slurry of calcium oxide or calcium carbonate to absorb SO_2 in a wet scrubber. The chemistry is quite complex, involving many side reactions. The overall reactions are those of SO_2 with lime (CaO) or limestone ($CaCO_3$) to form calcium sulfite ($CaSO_3$) with some oxidation of the sulfite to form calcium sulfate ($CaSO_4$). Alkaline fly ashes, if present, also contribute alkalinity in the form of soluble calcium, magnesium, and/or sodium oxides which participate in the SO_2 removal reactions.

The calcium sulfite and sulfate crystals precipitate in a reaction vessel or hold tank which is designed to provide adequate residence time for solids precipitation as well as for dissolution of the alkaline additive. The hold tank effluent is recycled to the scrubber to absorb additional SO_2. A slipstream from the hold tank is sent to a solid-liquid separator to remove the precipitated solids from the system. The waste solids are generally disposed of by ponding or landfill.

The basic design of a lime/limestone scrubbing system can be divided into the following process areas:

1) SO_2 absorption
2) Solids separation
3) Solids disposal

A simplified flow diagram is presented in Figure 2.2.1-1.

1) SO_2 absorption - Absorption of SO_2 takes place in a wet scrubber using lime or limestone in a circulating slurry. Particulates can be removed in the SO_2 absorber or ahead of the absorber by an electrostatic precipitator, wet scrubber, baghouse, or mechanical collector. The selection of a method for removal of particulates is based on economics and operational reliability. Removing particulates in the SO_2 absorber increases the solids load in the SO_2 scrubbing system.

After leaving the particulate removal device, the flue gas enters the wet SO_2 scrubber where absorption occurs. The overall reactions of gaseous SO_2 with the alkaline slurry yielding $CaSO_3 \cdot \frac{1}{2}H_2O$ are shown in Equations 2.2.1-1 and 2.2.1-2.

For lime systems:

$$SO_{2(g)} + CaO_{(s)} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O_{(s)} \qquad (2.2.1\text{-}1)$$

For limestone systems:

$$SO_{2(g)} + CaCO_{3(s)} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O_{(s)} + CO_{2(g)} \qquad (2.2.1\text{-}2)$$

The solid sulfite is only very slightly soluble in the scrubbing liquor and thus will precipitate to form an inert solid for disposal. In the lime system some CO_2 may also be absorbed from the flue gas and will react in a similar fashion to form solid calcium carbonate.

In most cases some oxygen will also be absorbed from the flue gas or surrounding atmosphere. This leads to oxidation of absorbed SO_2 and precipitation of solid $CaSO_4 \cdot 2H_2O$. The overall reaction for this step is as follows:

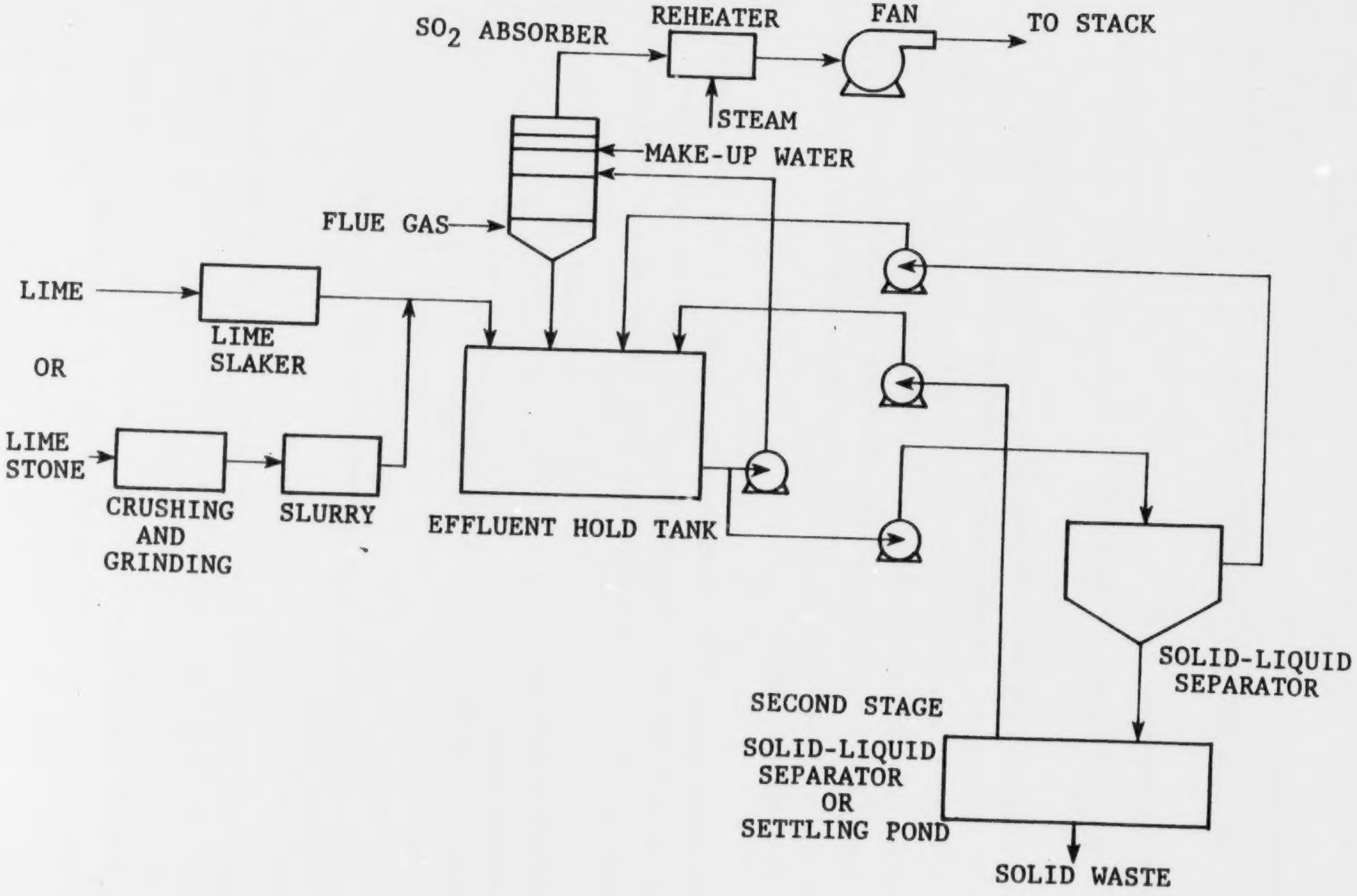


Figure 2.2.1-1. Process Flow Diagram Lime/Limestone Wet Scrubbing.

For lime systems:

$$SO_{2(g)} + \frac{1}{2}O_{2(g)} + CaO_{(s)} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O_{(s)} \qquad (2.2.1\text{-}3)$$

For limestone systems:

$$SO_{2(g)} + \frac{1}{2}O_{2(g)} + CaCO_{3(s)} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O_{(s)} + CO_{2(g)} \qquad (2.2.1\text{-}4)$$

The extent of oxidation can vary considerably, normally ranging anywhere from almost zero to 40 percent. In some systems treating dilute SO_2 flue gas streams, sulfite oxidations as high as 90 percent have been observed. The actual mechanism for sulfite oxidation is not completely understood, although the rate is known to be a strong function of oxygen concentration in the flue gas and liquor pH. It may also be increased by trace quantities of catalyst in fly ash entering the system.

Various types of gas-liquid contactors can be used as the SO_2 absorber. These differ in SO_2 removal efficiency as well as operating reliability. General types of contactors that have been used for SO_2 removal include:

- venturi scrubbers,
- packed towers,
- spray towers (horizontal and vertical),
- tray towers,
- grid towers, and
- mobile bed absorbers (such as marble bed and turbulent contact absorber (TCA)).

The liquid to gas ratio (L/G) generally varies widely depending mainly upon the type of contactor, the flue gas SO_2 concentration, and the required removal efficiency. A discussion of contactor types is found in Section 2.2.1.1D, Factors Affecting Performance. Simple impingement devices are placed downstream from the absorber to remove mist entrained in flue gas.

The effluent hold tank or reaction tank receives the lime or limestone feed slurry and absorber effluent. In addition, settling pond water and clarifier overflow can be sent to the hold tank. The tank is equipped with an agitator for uniform composition. The volume of the hold tank is sized to allow adequate residence time for calcium sulfite and sulfate precipitation and for lime or limestone dissolution. Too little residence time in the hold tank can cause nucleation of calcium sulfite and sulfate solids in the scrubber, resulting in scaling.

Design of the reaction tank is specific for each installation and is dependent upon the process precipitation rates and the lime or limestone dissolution rates. Bechtel reports that a liquid residence time of between 5-15 minutes has been tested at Shawnee and used in some full-scale systems.[6]

The feed material for a lime scrubbing process is usually produced by calcining limestone. Feed for a limestone process generally comes directly from the quarry and is then reduced in size by crushing and grinding. The lime or limestone is mixed with water to make a 25-60 percent solids slurry.

2) Solids separation - A continuous slurry stream of 10-15 percent solids is removed from the hold tank and recycled to the absorber. The flow rate of this stream is generally dependent on the quantity of SO_2 to be removed from the flue gas. A bleed stream is removed from this recycle stream and is dewatered to minimize the area needed for sludge disposal. Dewatering can be accomplished in a variety of ways depending on the location of the disposal site and the method of disposal used.

For systems with on-site ponding, the bleed stream may be pumped directly from the effluent hold tank to the pond. The supernatant liquour may then be recycled back to the hold tank. Further thickening of the sludge can be achieved depending on the physical properties of the sludge. A thickening device such as clarifier can be used to increase the solids content to about 30 percent. A vacuum filter or centrifuge could be used to raise the solids content to about 50 percent or higher. A combination of vacuum filtering and forced oxidation of the precipitated solids can increase the solids content to 85 percent or greater. Further details are given in Section 2.2.1.1D, Factors Affecting Performance.

3) Solids disposal - Sludge disposal is one of the main disadvantages of lime/limestone FGD systems in comparison to recovery processes. While most sludge disposal studies have been aimed at utility applications where the magnitude of sludge production imposes large area requirements as illustrated in Table 2.2.1-1, it might also be useful to identify the disposal procedure of an industrial user of lime/limestone stystem. Rickenbacker Air Force Base discharges its thickened sludge to a 5-acre lined pond located about 700 feet from the absorber.[7] The pond is approximately 450 feet long, 250 feet wide and 12 feet deep with a pond life expectancy of well over five years at a coal combustion rate of 200,000 tons of 5 percent sulfur coal. Estimates of sludge production rates for typical industrial boiler sized limestone FGD systems are given in Table 2.2.1-2.

Dewatered sludge is generally sent to a pond or landfill for disposal. If land is available on-site, a large pond is usually constructed to settle out the solids and reuse the supernatant liquor. Both of the industrial boiler facilities that use a lime/limestone FGD process dispose of their dewatered sludge in ponds. Commercial "stabilization" methods are currently in use at some utilities to convert the sludge to structurally stable landfill material with low permeability.[8,9] These methods could be used when on-site disposal is not possible. The stabilized material can then be trucked to an off-site area for landfill. The stabilized material can then be trucked to an off-site area for landfill. Similarly, combined disposal of ash and sludge in certain cases results in a structurally stable disposal material.[10] Other disposal alternatives currently under study by EPA include deep mine filling, strip mine filling and ocean dumping [11] They are being evaluated both from economic and environmental standpoints.

TABLE 2.2.1-1. ESTIMATED QUANTITY OF FLUE GAS DESULFURIZATION WASTES AND ASH FROM SELECTED COALS - 1000 MW_e PLANT [12]

	A	B	C
Ash Content of Coal (%)	15	8	15
Sulfur Content of Coal (%)	3.5	0.8	2.0
Heating Value of Coal (Btu/lb)	12,500	8,500	12,500
Annual Coal Use (million tph)	2.63	3.85	2.63
Sulfur Emission Standards	1.2 lb/10^6 Btu	50% removal	1.2 lb/10^6 Btu
Annual Sludge Production (tpy dry)	315,000	35,000	160,000
Annual Ash Production (tpy dry)	395,000	310,000	395,000
Annual Total Solid Wastes (tpy dry)	710,000	345,000	555,000

TABLE 2.2.1-2. ESTIMATED QUANTITY OF SLUDGE FROM INDUSTRIAL BOILER LIMESTONE FGD SYSTEMS

(Ash Free Basis, 90 Percent SO_2 Removal)

Boiler Size (10^6 BTU/hr)	Coal Sulfur Content (%)	Mass Production Rate		Volume Production	
		g/sec	(lb/hr)	cc/sec	(ft^3/hr)
30	3.5	90	(712)	69	(8.8)
30	0.6	21	(166)	14	(1.8)
75	3.5	227	(1796)	174	(22.1)
75	0.6	52	(414)	35	(4.4)
150	3.5	457	(3624)	352	(44.6)
150	0.6	105	(830)	69	(8.8)
200	3.5	607	(4812)	467	(59.3)
200	0.6	140	(1106)	92	(12)

B. Development status--Lime/limestone FGD systems have been commercially demonstrated at several utility and industrial boiler installations in the U.S., Japan, and several other countries. The wide use of lime/limestone FGD systems is due primarily to the fact that these systems are the most technically advanced and generally the most economically attractive, at least for utility situations.[13]

Lime/limestone scrubbing was first used to control SO_2 emissions on commercial boiler pilot plants in England about 40 years ago. This led to full-scale gas washing plants which proved successful in removing SO_2 and dust from stack gas. The lime/limestone process was also the first SO_2 stack gas scrubbing technology used in this country, mainly due to the fact that there was more experience behind the process and that it produced a throwaway sludge rather than a marketable by-product which could require a considerable marketing effort in many instances to dispose of. The trend to using lime/limestone scrubbing for SO_2 removal is strong today due to rapid progress in coping with the many process problems and a clearer understanding of process economics.

1) Commercial applications - The commercial use of lime/limestone systems for coal-fired boilers is addressed in three categories:

- Utility boilers in the U.S.
- Industrial boilers in the U.S.
- Foreign applications

Utility installations - Lime/limestone systmes are the most widely used form of FGD selected by the U.S. electric utility industry. As of March, 1978, some 34,000 MW of coal-fired electrical generating capacity in the United States had been commited to lime/limestone scrubbing. This figure includes 28 facilities in operation, 35 under construction, and another 16 in the planning stages (*i.e.*, contract awarded, letter of intent signed, or requesting/evaluating bids).

Wide use of lime/limestone scrubbing is true for both new and retrofit cases. As shown in Table 2.2.1-3, 94 percent of the total MW of controlled capacity of new systems currently operational is by lime/limestone processes. For operational retrofit situations, 84 percent of the total capacity is controlled by lime/limestone. Data are also summarized for units under construction and in the planning stages. Operating lime/limestone scrubbing units on United States power plants are shown in Table 2.2.1-4, units under construction in Table 2.2.1-5, and planned units in Table 2.2.1-6. It might also be noted that an additional 11 lime/limestone systems which were operational within the last five years are currently in a "terminated" state for one reason or another.

Industrial installations - Industrial usage of lime/limestone FGD systems in the U.S. has been quite limited. As of January 1979 there were two operational and two planned lime/limestone system with a combined capacity of 489,000 scfm. This represents less than 5 percent of the total controlled (operational, under construction, and planned) U.S. industrial boiler capacity. The lime/limestone systems are summarized in Table 2.2.1-7.

One of the most significant lime/limestone applications is the Bahco system installed at the Rickenbacker Air Force Base (RAFB) in Columbus, Ohio. he RAFB installation is the first Bahco system installed in the United States. It is designed to treat flue gases from an equivalent coal-firing rate of 58.6 MW_t (200×10^6 Btu/hr) under full load conditions as well as flue gases from summer load conditions, about 5.8 MW_t (20×10^6 Btu/hr). EPA has recently completed a two year test program on this system. Although no continuous monitoring data are available from these tests, the discrete data sets from the various factorial runs showed the system to consistently exceed design specifications. Data on process operations for this system will be presented in the appropriate sections of this chapter.[19]

TABLE 2.2.1-3. SUMMARY OF NEW AND RETROFIT FGD SYSTEMS FOR U.S. UTILITY INDUSTRY BY PROCESS[14]

Process	New or retrofit	Operational		Construction		Contract awarded		Planned		Total No. of plants	
		No.	MW	No.	MW	No.	MW	No.	MW	No.	MW
Lime	N	4	2450	10	4565	2	1425	0	0	16	8440
	R	8	1650	0	0	0	0	2	660	10	2310
Lime/Alkaline Fly Ash	N	3	1170	1	500	2	1400	1	527	7	3597
	R	0	0	0	0	0	0	3	579	3	579
Lime/Limestone	N	0	0	0	0	0	0	0	0	0	0
	R	2	20	0	0	0	0	0	0	2	20
Limestone	N	8	4443	23	9620	9	4783	5	2880	45	21726
	R	3	790	1	425	1	575	0	0	5	1790
Subtotal - Lime/Limestone	N	15.	8063.	34.	14685.	13.	7608.	6.	3407.	68.	33763.
	R	13.	2460.	1.	425.	1.	575.	5.	1239.	20.	4699.
Aqueous Carbonate	N	0	0	0	0	0	0	0	0	0	0
	R	0	0	0	0	1	100	0	0	1	100
Aqueous Carbonate/Fab. Filter	N	0	0	0	0	1	400	0	0	1	400
	R	0	0	0	0	0	0	0	0	0	0
Double Alkali	N	0	0	2	825	0	0	0	0	2	825
	R	0	0	1	277	0	0	0	0	1	277
Magnesium Oxide	N	0	0	0	0	0	0	0	0	0	0
	R	1	120	0	0	0	0	3	726	4	846
Not Selected	N	0	0	0	0	1	300	18	9500	19	9800
	R	0	0	0	0	0	0	4	2100	4	2100
Regenerable Not Selected	N	0	0	0	0	0	0	0	0	0	0
	R	0	0	0	0	0	0	1	650	1	650
Sodium Carbonate	N	1	125	1	509	0	0	1	125	3	759
	R	2	250	0	0	0	0	0	0	2	250
Wellman Lord	N	0	0	1	500	0	0	1	500	2	1000
	R	0	0	1	180	0	0	0	0	1	180
Wellman Lord/Allied Chemical	N	1	375	0	0	0	0	0	0	1	375
	R	1	115	1	340	0	0	0	0	2	455
Totals	N	17.	8563.	38.	16519.	15.	8308.	26.	13532.	96.	46922.
	R	17.	2945.	4.	1222.	2.	675.	13.	4715.	36.	9557.
Lime/Limestone % of Total MW	N	94		89		92		25		72	
	R	84		35		85		26		49	

TABLE 2.2.1-4. SUMMARY OF OPERATIONAL LIME/LIMESTONE FGD SYSTEMS FOR U.S. UTILITIES AS OF MARCH 1978[15]

Process	Utility company/power station	New or retrofit	Size of FGD unit (MW)	Vendor	Start up mo/yr	Years Operational
Lime	Columbus & Southern Ohio Elec. Conesville No. 5	N	400	UOP	1/77	1
	Columbus & Southern Ohio Elec. Conesville No. 6	N	400	UOP	4/78	½
	Duquesne Light Phillips Power Station	R	410	Chemico	7/73	5
	Duquesne Light Elrama Power Station	R	510	Chemico	10/75	3
	Kansas City Power & Light Hawthorn No. 3	R	140	Combustion Engineering	11/72	6
	Kansas City Power & Light Hawthorn No. 4	R	100	Combustion Engineering	8/72	6
	Kentucky Utilities Green River, Units 1, 2, and 3	R	64	American Air Filter	9/75	3
	Louisville Gas & Electric Paddy's Run No. 6	R	65	Combustion Engineering	4/73	5½
	Louisville Gas & Electric Cane Run No. 4	R	178	American Air Filter	8/76	2
	Louisville Gas & Electric Cane Run No. 5	R	183	Combustion Engineering	12/77	1
	Minnkota Power Coop. Milton R. Young No. 2	N	450	ADL/Combustion Equipment Assoc.	9/77	1
	Montana Power Company Colstrip No. 2	N	360	ADL/Combustion Equipment Assoc.	7/76	2

TABLE 2.2.1-4. (Continued)

Process	Utility company/power station	New or retrofit	Size of FGD unit (MW)	Vendor	Start up mo/yr	Years operational
Lime (Cont'd)	Montana Power Company Colstrip No. 1	N	360	ADL/Combustion Equipment Assoc.	11/75	3
	Pennsylvania Power Co. Bruce Mansfield No. 1	N	825	Chemico	4/76	2½
	Pennsylvania Power Co. Bruce Mansfield No. 2	N	825	Chemico	7/77	1
Limestone	Arizona Public Service Cholla No. 1	R	115	Research-Cottrell	10/73	5
	Indianapolis Power & Light Co. Petersburg No. 3	N	530	UOP	10/77	1
	Kansas City Power & Light LaCygne No. 1	N	820	Babcock & Wilcox	2/73	5½
	Kansas Power & Light Lawrence No. 4	R	125	Combustion Engineering	12/68	10
	Kansas Power & Light Lawrence No. 5	N	400	Combustion Engineering	11/71	7
	Northern States Power Co. Sherburne County Station No. 1	N	710	Combustion Engineering	3/76	2½
	Northern States Power Co. Sherburne County Station No. 2	N	710	Combustion Engineering	4/77	1½
	South Carolina Public Service Winyah No. 2	N	280	Babcock & Wilcox	7/77	1
	Springfield City Utilities Southwest No. 1	N	200	UOP	4/77	1½

TABLE 2.2.1-4. (Continued)

Process	Utility company/power station	New or retrofit	Size of FGD unit (MW)	Vendor	Start up mo/yr	Years operational
Limestone (Cont'd)	Tennessee Valley Authority Widows Creek No. 8	R	550	TVA	5/77	1
	Texas Utilities Co. Martin Lake No. 1	N	793	Research Cottrell	8/77	1
Lime/ Limestone	Tennessee Valley Authority Shawnee No. 10B	R	10	Chemico	4/72	6½
	Tennessee Valley Authority Shawnee No. 10A	R	10	UOP	4/72	6½

TABLE 2.2.1-5. SUMMARY OF LIME/LIMESTONE SYSTEMS UNDER CONSTRUCTION FOR U.S. UTILITIES AS OF MARCH 1978[16]

Process	Utility company/power station	New or retrofit	Size of FGD unit (MW)	Vendor	Start up mo/yr
Lime	Allegheny Power System Pleasants No. 1	N	625	Babcock & Wilcox	3/79
	Allegheny Power System Pleasants No. 2	N	625	Babcock & Wilcox	3/80
	Big Rivers Electric Coop Corp. Reid Steam Station No. 2	N	250	American Air Filter	12/79
	Big Rivers Electric Coop Corp. Reid Steam Station No. 3	N	240	American Air Filter	12/80
	Cooperative Power Association Coal Creek No. 1	N	545	Combustion Engineering	2/79
	Cooperative Power Association Coal Creek No. 2	N	545	Combustion Engineering	11/79
	Louisville Gas & Electric Mill Creek No. 3	N	425	American Air Filter	7/78
	Louisville Gas & Electric Mill Creek No. 4	N	495	American Air Filter	6/80
	Minnesota Power & Light Co. Clay Boswell No. 4	N	500	Peabody Engineering	5/80
	Utah Power & Light Co. Emery No. 1	N	400	Chemico	12/78
	Utah Power & Light Co. Huntington No. 1	N	415	Chemico	5/78
Limestone	Alabama Electric Coop Tombigbee No. 2	N	225	Peabody Engineering	6/78

TABLE 2.2.1-5. (Continued)

Process	Utility company/power station	New or retrofit	Size of FGD unit (MW)	Vendor	Start up mo/yr
Limestone (Cont'd)	Alabama Electric Coop Tombigbee No. 3	N	225	Peabody Engineering	6/79
	Arizona Electric Power Coop Apache No. 3	N	200	Research Cottrell	4/79
	Arizona Electric Power Coop Apache No. 2	N	200	Research Cottrell	6/78
	Arizona Public Service Cholla No. 2	N	250	Research Cottrell	6/78
	Basin Electric Power Coop Laramie River No. 1	N	550	Research Cottrell	4/80
	Basin Electric Power Coop Laramie River No. 2	N	550	Research Cottrell	10/80
	Board of Municipal Utilities Sikeston Power Station	N	235	Babcock & Wilcox	6/81
	Brazos Electric Power Coop San Miguel No. 1	N	400	Babcock & Wilcox	6/80
	Central Illinois Light Co. Duck Creek No. 1	N	400	Riley Stocker/Environeering	8/78
	Colorado Ute Electric Assn. Craig No. 1	N	450	Peabody Engineering	3/79
	Colorado Ute Electric Assn. Craig No. 2	N	450	Peabody Engineering	3/79
	Commonwealth Edison Powerton No. 51	R	425	UOP	12/79

TABLE 2.2.1-5. (Continued)

Process	Utility company/power station	New or retrofit	Size of FGD unit (MW)	Vendor	Start up mo/yr
Limestone (Cont'd)	Kansas Power & Light Jeffery No. 1	N	680	Combustion Engineering	6/78
	Kansas Power & Light Jeffery No. 2	N	680	Combustion Engineering	6/80
	Salt River Project Coronado No. 1	N	350	Pullman Kellogg	4/79
	Salt River Project Coronado No. 2	N	350	Pullman Kellogg	4/80
	Southern Illinois Power Coop Marion No. 4	N	184	Babcock & Wilcox	6/78
	Southern Mississippi Electric R. D. Morrow No. 1	N	180	Riley Stoker/Environeering	5/78
	Southern Mississippi Electric R. D. Morrow No. 2	N	180	Riley Stoker/Environeering	8/78
	Texas Power and Light Co. Sandow No. 4	N	545	Combustion Engineering	7/80
	Texas Utilities Co. Martin Lake No. 2	N	793	Research Cottrell	2/78
	Texas Utilities Co. Martin Lake No. 3	N	793	Research Cottrell	12/78
	Texas Utilities Co. Monticello No. 3	N	750	Chemico	2/78

TABLE 2.2.1-6. SUMMARY OF PLANNED LIME/LIMESTONE FGD SYSTEMS AS OF 03/78[17]

Process	Utility company/power station	New of retrofit	Size of FGD unit (MW)	Vendor	Start up mo/yr
	Contract Awarded				
Limestone	Arizona Public Service Cholla No. 4		350	Research Cottrell	6/80
Lime	Cincinnati Gas & Electric Co. East Bend No. 2	N	600	Babcock & Wilcox	1/81
Limestone	Hoosier Cooperative Merom No. 1	N	490	Not selected	12/80
Limestone	Hoosier Cooperative Merom No. 2	N	490	Not selected	10/81
Limestone	Lakeland Utilities McIntosh Power Plant Unit No. 3	N	350	Not selected	10/81
Lime/ Alkaline Fly Ash	Montana Power Co. Colstrip No. 3	N	700	ADL/Combustion Equip. Associate	7/80
Lime/ Alkaline Fly Ash	Montana Power Co. Colstrip No. 4	N	700	ADL/Combustion Equip. Associate	7/81
Limestone	Northern States Power Co. Sherburne County Station No. 3	N	860	Combustion Engineering	5/81
Limestone	Northern States Power Co. Sherburne County Station No. 4	N	860	Combustion Engineering	5/83
Lime	Pennsylvania Power Co. Bruce Mansfield No. 3	N	825	Pullman Kellogg	4/80
Limestone	Springfield Water Light & Power Dallman No. 3	N	190	Research Cottrell	7/80

TABLE 2.2.1-6. (Continued)

Process	Utility company/power station	New or retrofit	Size of FGD unit (MW)	Vendor	Start up mo/yr
Limestone	Tennessee Valley Authority Widows Creek No. 7	R	575	Combustion Engineering	
Limestone	Texas Municipal Power Agency Gibbons Creek Unit No. 1	N	400	Combustion Engineering	1/82
Limestone	Texas Utilities Co. Martin Lake No. 4	N	793	Research Cottrell	11/82
	Letter of Intent Signed				
Lime/ Alkaline Fly Ash	Wisconsin Power & Light Co. Columbia No. 2	N	527	Chemico	1/80
	Requesting/Evaluating Bids				
Limestone	Indianapolis Power & Light Co. Petersburg No. 4	N	530	Not Selected	4/82

TABLE 2.2.1-7. SUMMARY OF COMMITTED LIME/LIMESTONE SYSTEMS FOR U.S. INDUSTRIAL BOILERS AS OF MARCH 1978[18]

Process	Vendor	Company/ Location	New or retrofit	Size of FGD unit		characteristics	Startup mo/yr
				scfm	MW		
Lime	Koch Engineering	Armco Steel Middletown, OH	R	84,000	42	Coal 0.8% sulfur	1975
Lime	Carborundum Environ. Systems, Ltd.	Carborundum Abrasives Buffalo, NY	R	30,000	15	Coal 2.2% sulfur	1980
Lime/ Limestone	Research Cottrell-Bahco	Rickenbacker Air Force Base Columbus, OH	R	55,000	27	Coal 3.6% sulfur	1976
Limestone		Dupont Texas	N	320,000	160	Coal 0.5% sulfur	1982

Foreign installations - Foreign applications of lime/limestone scrubbing also serve to illustrate its commercial availability. The most widespread use of this technology is in Japan, with a few facilities also in Germany, Sweden, Russia, and England.

Japanese wet lime/limestone FGD installations by the end of 1977 totalled 94 plants with a combined flue gas capacity of 43.4×10^6 Nm^3/hr, equivalent to about 13,000 MW_e or about 50% of all Japanese FGD capacity. This total includes industrial and utility boilers, sintering plants and sulfuric acid plants.[20] According to a report for EPA prepared by Dr. Jumpei Ando, lime/limestone FGD applications are distributed as shown in Table 2.2.1-8. The emphasis in Japan is on oil-fired boilers. Practically all the processes make salable gypsum, and "operabilities" are reported to exceed 95 percent.[22]

2) Recent Improvements - Lime/limestone FGD technology, as summarized in this report, can be regarded as established and commercially available. However, attempts are constantly being made to improve the overall performance of the technology and to increase its acceptability. Current research in the areas of mass transfer additives and cocurrent scrubbing offer the potential for improved process performance and cost savings, and ongoing work in the area of forced oxidation has the potential for improving the process solid waste characteristics thus increasing the overall acceptability of the technology. Each of these areas of recent developments is briefly described in this section.

Mass transfer additives - Additives such as MgO, Na_2CO_3, and organic acids can be used to improve mass transfer in lime/limestone systems, thus improving SO_2 removal, increasing limestone utilization, and permitting the use of

TABLE 2.2.1-8. SUMMARY OF JAPANESE LIME/LIMESTONE INSTALLATIONS[21]

Type Installation	Fuel	Capacity $10^6 Nm^3/hr$	Capacity MWe	Percent of Total Lime/ Limestone Capacity
Utility Boiler	Oil	24.1	7530	55.5
Utility Boiler	Coal	3.6	1125	8.3
Industrial Boiler	Oil	3.7	1150	8.6
Industrial Boiler	Coal	0.8	250	1.8
Chemical Plant	-	11.2	3500	25.8
	Total	43.4	13555	100

more simple scrubber types (such as cocurrent scrubbers). SO_2 removal can be increased from 85 to 95 percent through use of these additives.[23] The improvement in limestone utilization also improves process operability, since excess limestone (which can cause plugging of the scrubber mist eliminator) is minimized. The use of additive also enhances the effects of dissolved alkalinity by increasing the concentrations of HCO_3^- and $SO_3^=$ which react with SO_2 as shown below:

$$HCO_3^- + SO_2 \rightarrow CO_2 + HSO_3^- \qquad 2.2.1\text{-}5$$

$$SO_3^= + SO_2 + H_2O \rightarrow 2\ HSO_3^- \qquad 2.2.1\text{-}6$$

Alkali additives, MgO, Na_2CO_3, and NH_3, have received the most attention to date. MgO has been tested extensively with lime and limestone scrubbing in both pilot and full scale systems. Na_2CO_3 and NH_3 have been used in double alkali processes and have been patented as additives for limestone-slurry scrubbers. The effects of Na^+ and Mg^{2+} addition on total dissolved alkalinity and improved scrubber performance have been calculated and reported by Rochelle.[24] It is believed that NH_3^+ probably behaves similarly to Na^+. When the soluble alkali carbonates or oxides are added to a slurry scrubbing system they accumulate in solution primarily as the soluble sulfate.

Chloride concentrations can reduce the apparent effectiveness of alkali additives by reacting with the additive to form inactive chloride salts. For a given alkali concentration, dissolved chloride reduces accumulated sulfate ion by 1 mole of sulfate per 2 moles of chloride. Consequently, there is little effect on dissolved alkalinity until an amount of alkali has been added equivalent to the chloride concentrations. In some scrubbing installations the alkali requirement will be dictated by the chloride level and most of the alkali will be present as the inactive chloride.

Buffer additives are also being evaluated for their effectiveness in enhancing the liquid phase mass transfer of limestone scrubbing systems. Several carboxylic acids have been proposed by Rochelle for this application[25], and recent testing by EPA has demonstrated the effectiveness of adipic acid as a buffer additive.[26]

Recent results of EPA's Shawnee Test Facility, reported at the fifth FGD Symposium indicated that adipic acid concentrations of 700 to 1500 ppm increased SO_2 removals of the system to greater than 90 percent while decreasing the annual revenue requirements by 0.3 to 0.4 mill/kwh. Although this system has experienced decomposition of the adipic acid additive, the overall economics are still favorable. Long-term effects of organic buffers on sludge formation and stability are still under investigation.[27]

Concurrent Scrubbing - Cocurrent scrubbers differ from standard counter current scrubbers in that both the gas to be scrubbed and the scrubbing liquor enter the top of the scrubber and flow in the same direction through the absorber. Advantages of this type scrubber are that higher gas velocities and lower pressure drops occur than the conventional countercurrent scrubbers which can result in lower capital and operating costs. Flue gas from the bottom of the scrubber experiences a change in direction to "knock-out" the majority of entrained scrubbing liquor before going to a mist eliminator. The actual scrubber may use sprays or packing for gas/liquid contact.

Higher gas velocities through the scrubber are possible; therefore, smaller diameter absorbers may be used. Since spray towers are generally designed for length to diameter (L/D) ratios of about 2 for good gas distribution, cocurrent operations are attractive from a liquid pumping height standpoint. Reduced pressure drop will result in fan cost savings and in lower energy requirements and operating costs.

In addition, the cocurrent scrubber concept should reduce the effect of plugged nozzles on poor gas distribution. In countercurrent designs the gas tends to flow where the pressure drop is lowest (*i.e.*, in the area of a plugged nozzle) which results in reduced mass transfer. However, in the cocurrent scrubber, the liquid flow imparts additional energy to the gas so gas will not tend to flow toward plugged nozzles. Instead, the gas flow should be higher in regions where the liquid flow is higher.

Forced oxidation - Forced oxidation systems are presently being marketed although further development work is being performed by EPA and others. As the name implies, forced oxidation systems increase the amount of sulfite which is oxidized to sulfate. This is generally done by sparging air into the hold tank. Potential advantages include improved waste solid characteristics and better limestone utilizations. If oxidations approaching 100 percent are achieved, the gypsum waste solid may be suitable for wallboard production.[28]

Gypsum solids are more easily dewatered than sludges with a high concentration of calcium sulfite since the gypsum solids are larger, denser, and have a faster settling rate. Gypsum solids have been shown to be suitable for landfill disposal. A liquor purge may be necessary to control soluble species in systems that have waste products with a high solids content, but the purge would be no larger than the amount of liquor discharged with the waste solids from a system that does not use forced oxidation.[29]

One disadvantage of forced oxidation is the elimination of $CaSO_3$ solids so that they are not available for mass transfer. The use of additives to enhance mass transfer may thus be attractive when used in conjunction with forced oxidation. Since forced oxidation also minimizes additive losses in the waste solids, forced oxidation and mass transfer additives appear to be particularly attractive when used together.[30]

C. Applicability--A major factor to be considered when assessing a lime/limestone system's applicability is available land. The area required for sludge disposal can be especially important when the industrial plant is located in a densely populated area such as the East Coast. Estimates of sludge disposal requirements were presented previously in Section 2.2.1.1.C.

In some cases the plant might be required to pay for off-site transportation and disposal. Disposal problems and costs can be limiting factors for both new and retrofit applications.

Sorbent availability should not limit process applicability since limestone is readily available material as is reflected by its low price of approximately $7/ton.[31] Sorbent price, however, will be dependent on transportation costs. Production of limestone in the United States was approximately 630 million tons in 1970. Since limestone is mined in many areas of the U.S., it will be reasonably close to many industrial boiler installations.

Lime is produced by calcining high-calcium limestone in kilns at about 1800-2300°F to drive off CO_2 and form CaO. The heat of reaction required to convert $CaCO_3$ to CaO is about 2.8×10^6 Btu/ton of CaO. In practice, heat input may vary from 4 to 10×10^6 Btu/ton of lime. Total lime consumption by the United States in 1975 (including limes processed within captive facilities) was 19.2 million tons.[32] Over 80 percent of the lime used in the United States is by chemical and industrial manufacturing, mostly as quicklime. Since limestone is used to make lime and is readily available, lime also can be obtained fairly easily. The cost of lime is presently about $40/ton [33] with a large portion of the cost being due to the fuel costs associated with lime production.

As presented in the preceding section, there are considerably more applications of lime/limestone scrubbing in the utility sector than in the industrial sector both in this country and abroad. However, the process is applicable to small boiler operations as well. Considerations affecting retrofittability and by-product utilization for industrial applications are similar to those for utility applications.

D. Factors affecting performance--

1) Design and operating considerations - Although lime/limestone scrubbing systems tend to be somewhat simple mechanically, they are more complex from a chemical process point of view. The removal of SO_2 from flue gas in a lime/limestone scrubbing process is a gas-liquid-solid mass trans-

fer phenomenon. First, sulfur dioxide must be transferred from the gas phase to the liquid phase in a scrubber. Then the sulfur species must be precipitated from the scrubbing liquor as insoluble calcium salts and disposed of.

Most lime/limestone systems tend to be at least partially liquid film mass transfer limited. Only at low SO_2 gas loadings do these systems approach gas film limiting situations. Liquid film resistance can be reduced by increasing the amount of liquid phase alkalinity available in the scrubber. As a rule, a large portion of the alkalinity required for SO_2 removal is derived from solids dissolution in the scrubber. Since solid-liquid reactions tend to be significantly slower than do liquid-liquid reactions, it is advantageous to minimize the amount of solids dissolution required by maximizing the soluble liquid phase alkalinity in the scrubber feed liquor. This is why systems which operate with high magnesium and sodium concentrations but low chloride levels exhibit higher SO_2 removals than systems which are lower in soluble alkalinity.[34]

Gas distribution can also be a major problem, particularly in large commercial units. Poor gas distribution will minimize the effective interfacial mass transfer area. Portions of the scrubber can become liquid phase alkalinity limited even though the total alkalinity entering the scrubber is sufficient for good SO_2 removal. Thus, the potential impact of gas distribution problems should be seriously considered in both process design and analysis.

In addition to mass transfer and gas distribution, several other design and operating variables should be considered in the design of a lime/limestone FGD system. The effect of the following variables, both on SO_2 absorption efficiency and on overall process operations are summarized in this section:

- L/G Ratio
- Slurry pH
- Inlet SO_2 Concentration

- Reaction Tank Design
- Effects of Soluble Species
- Number of Contact Stages
- Ash Removal
- Oxidation and Scale Prevention
- Water Balance
- Reheat
- Sludge Dewatering and Disposal
- Erosion and Corrosion

L/G ratio - For both lime and limestone systems higher SO_2 removal efficiencies are achieved at greater L/G ratios up to the point where flooding and poor gas distribution occurs. This trend is documented by performance data from several different scrubber operations. Data from the Shawnee test site which represents an eastern high-sulfur coal (*i.e.*, high inlet SO_2 concentration) is shown in Figures 2.2.1-2 through 2.2.1-5. Results from a western plant are presented in Figures 2.2.1-6 and 2.2.1-7. While these data represent numerous scrubber types and operating conditions as well as lime and limestone systems, the increasing trend of SO_2 removal with increasing L/G is evident in all.

Higher L/G's can be obtained by either increasing the liquor flow or reducing the gas flow through the scrubber. The first approach entails more pumping capacity, pipes, and slurry. The latter approach requires additional scrubbing capacity. The first method, illustrated in Figures 2.2.1-2 through 2.2.1-7, is more commonly used by designers for cost reasons. Shawnee spray tower data illustrate the effects of decreasing the gas velocity (Figure 2.2.1-8). Decreasing the gas velocity is much less effective for a packed or turbulent type contactor than a spray tower because increased agitation and mixing offset the decreased L/G ratio for these contactors.

Slurry pH - At higher pH levels (*i.e.*, greater lime or limestone feed stoichiometry), increased SO_2 removal efficiency occurs because more alkali is available and higher dissolution rates are achieved. However, scaling will occur if the pH becomes too high. Therefore, careful pH monitoring and control is

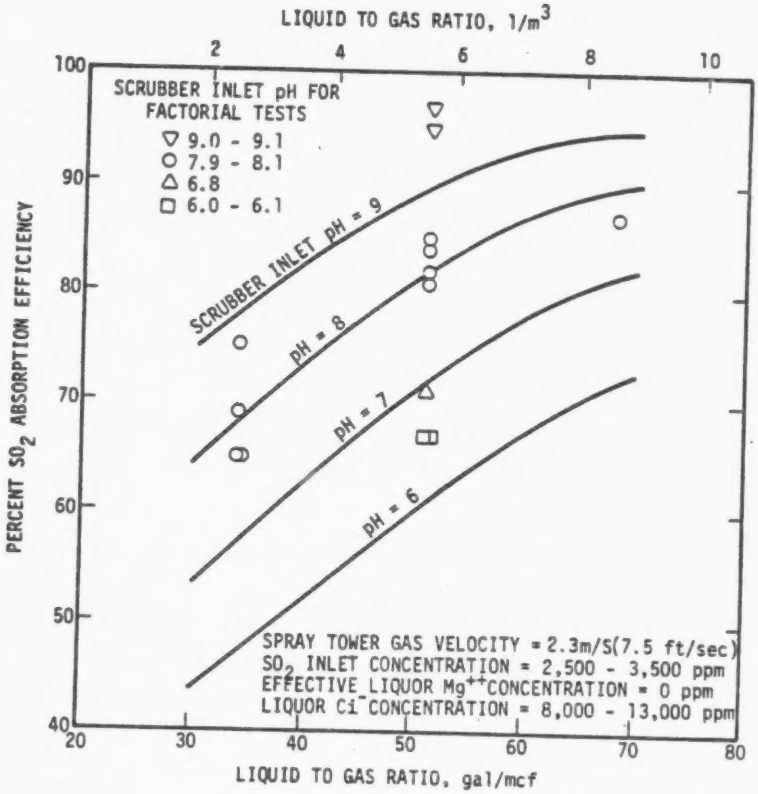


Figure 2.2.1-2. Liquid-to-gas ratio and scrubber inlet pH *versus* predicted and measured SO_2 removal, spray tower with lime, Shawnee plant.[35]

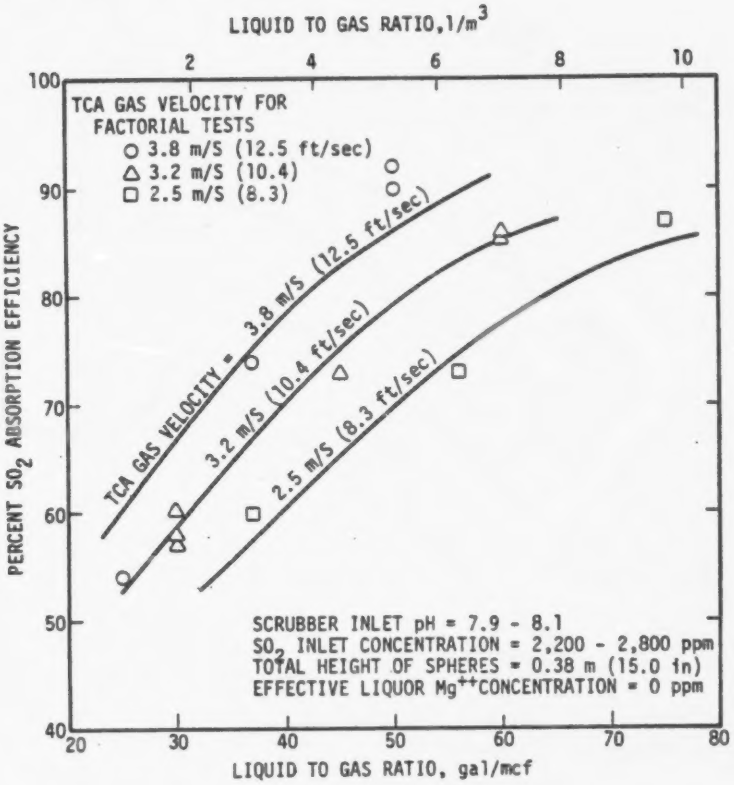


Figure 2.2.1-3. Liquid-to-gas ratio and scrubber gas velocity *versus* predicted and measured SO_2 removal, TCA with lime, Shawnee plant.[36]

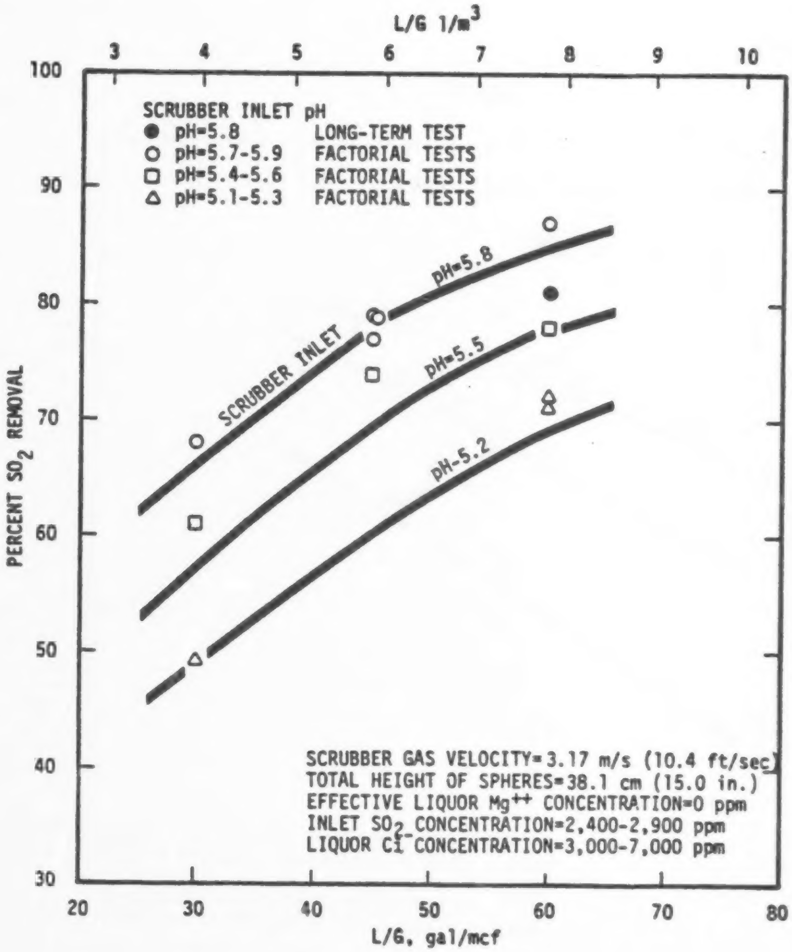


Figure 2.2.1-4. L/G ratio and scrubber inlet pH *versus* predicted and measured SO_2 removal - TCA with limestone - Shawnee plant.[37]

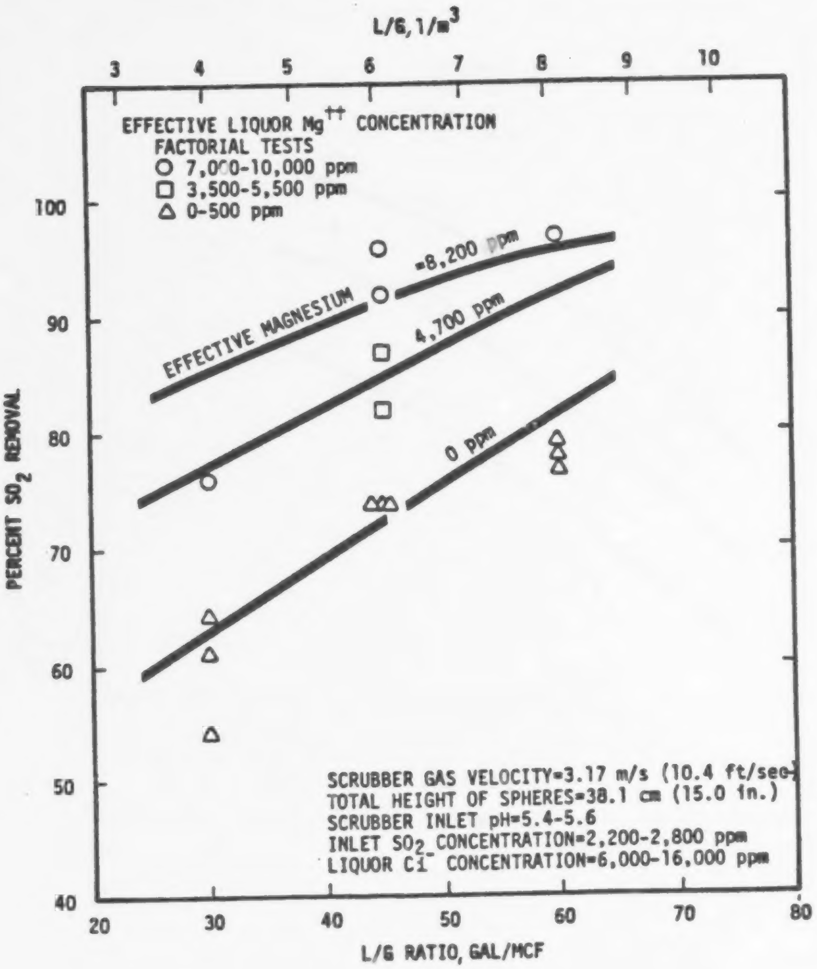


Figure 2.2.1-5. L/G ratio *versus* percent SO_2 removal at various magnesium ion concentrations TCA with limestone - Shawnee plant.[38]

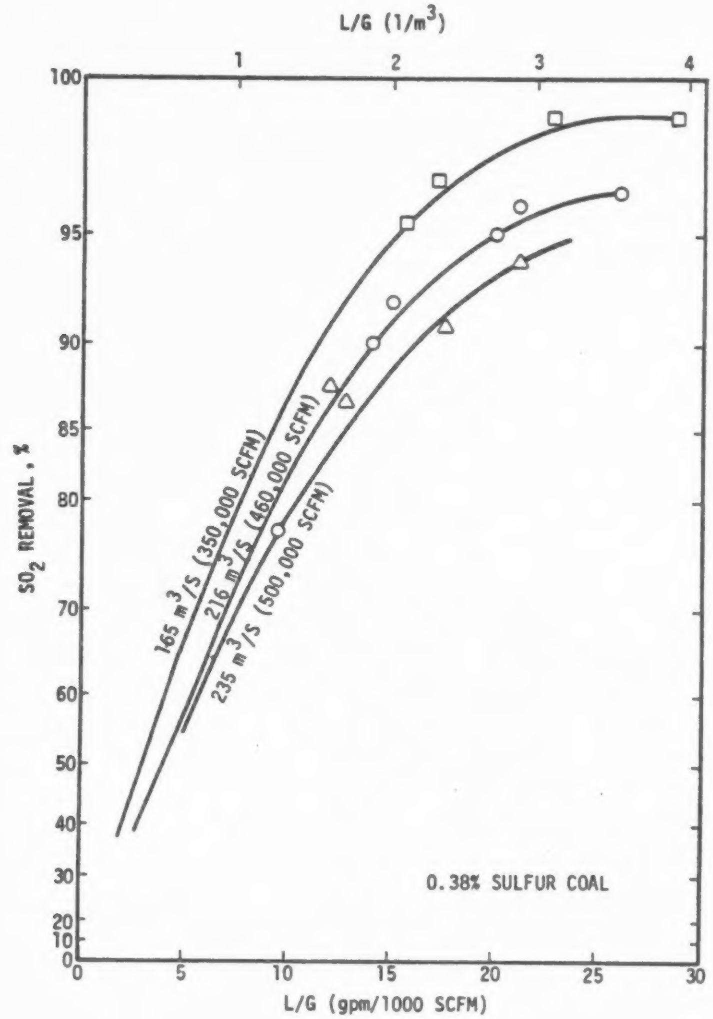


Figure 2.2.1-6. SO$_2$ removal *versus* L/G ratio, 170-MW horizontal module, lime sorbent, Mohave plant.[99]

[L/G's are for 1-stage of a 4-stage scrubber.
Total L/G's are 4 times the indicated value.]

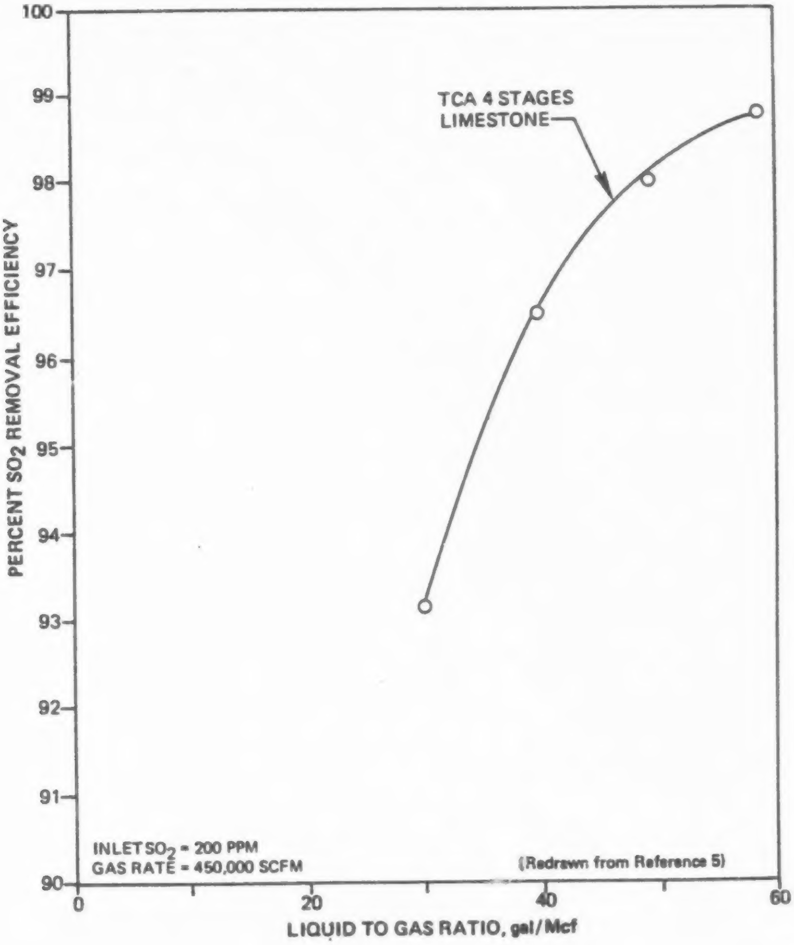


Figure 2.2.1-7. Effect of liquid-to-gas ratio on SO_2 removal efficiency with low sulfur coal at the Mohave power station.[40]

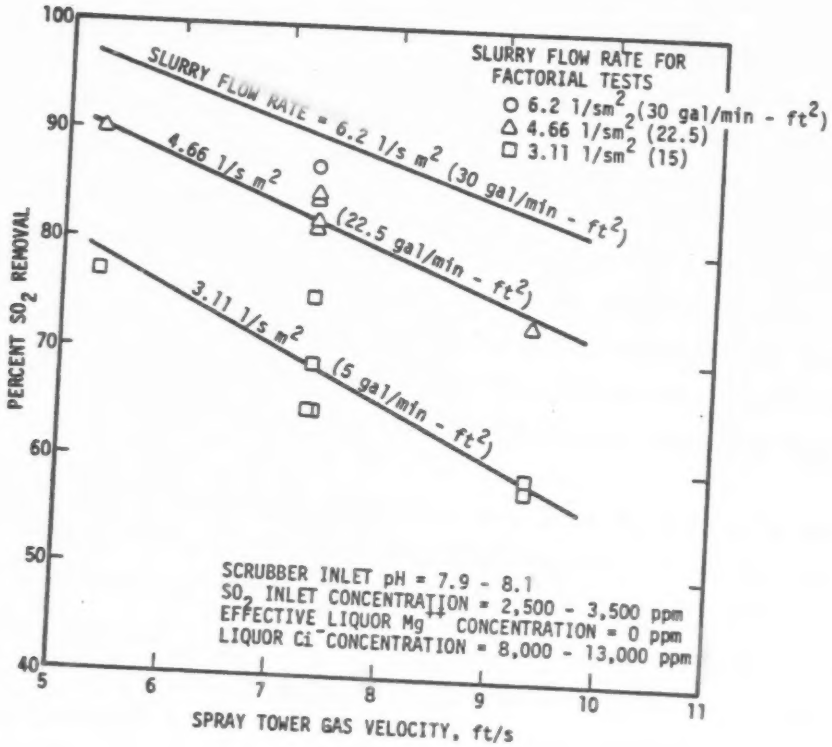


Figure 2.2.1-8. Gas velocity and slurry flow rate *versus* predicted and measured SO_2 removal, spray tower with lime, Shawnee plant.[41]

needed to optimize SO_2 removal, avoid scaling, and prevent low utilization of excess reagent.

Lime is much more reactive than limestone. In most cases, complete dissolution of lime occurs within 20-30 seconds of the time that the lime is introduced into the hold tank. Since lime is not a buffered alkali, higher pHs are maintained in lime systems compared to limestone systems. However, the combination of reduced reaction time and increased pH can lead to $Ca(OH)_2$ solids entering the scrubber in the feed slurry. This should be avoided since the dissolution of these lime solids will occur in the low pH environment of the scrubber. This, in turn, will result in locally high pHs and high relative saturations of calcium sulfite and sulfate. These conditions can lead to scale deposition on scrubber internals.

If the hold tank pH in a lime system is maintained below about 8, lime utilizations in the 95 percent range are possible.[42] Lime systems operating in the optimal pH range for high lime utilization can operate subsaturated with respect to calcium sulfite in the scrubber. In fact, since calcium sulfite dissolution is normally noted in lime scrubbers, it can be concluded that calcium sulfite scaling is not likely to be a problem, except during upset conditions.

Lowering a limestone system's slurry pH will increase the limestone dissolution rate and improve the utilization efficiency, but the resulting pH drop can lead to decreased SO_2 removal efficiencies.[43,44] An explanation for this is found by considering the dissolution step for $CaCO_3$.

$$CaCO_{3(s)} \rightleftarrows CaCO_{3(aq)} \rightleftarrows Ca^{++}_{(aq)} + CO^{=}_{3(aq)} \qquad (2.2.1\text{-}11)$$

The carbonate ion can then be reacted as follows:

$$CO_3^{=} + H^{+} \rightleftarrows HCO_3^{-} \qquad (2.2.1\text{-}12)$$

$$HCO_3^{-} + H^{+} \rightleftarrows H_2CO_3 \qquad (2.2.1\text{-}13)$$

The slurry pH in a limestone system remains buffered according to these equations. This accounts for the fact that the slurry pH is relatively insensitive to changes in the amount of dissolved limestone (and therefore dissolved alkalinity). Because large changes in liquid phase alkalinity can occur over fairly narrow pH shifts, SO_2 removal is generally a very sensitive function of the system pH in a liquid film limiting situation.

This buffering capacity of limestone also means that to achieve higher scrubber inlet pH's, increasingly larger amounts of limestone must be added to the hold tank. This results in larger amounts of limestone remaining undissolved in the hold tank effluent and higher limestone solids concentrations in the scrubber feed liquor. The amount of solid $CaCO_3$ available for reaction in the scrubber can have a major effect on SO_2 removal. As the concentration of $CaCO_3$ increases, generally, the surface area also increases. This can lead to increased dissolution. From this standpoint, an increase in scrubber pH in a limestone system will increase the removal for two reasons: 1) because it increases the amount of dissolved liquid phase alkalinity and 2) because it increases the amount of limestone solids in the circulating slurry. Attention must be given to the amount of $CaCO_3$ which dissolves in the scrubber since locally high relative saturations in the vicinity of dissolving limestone particles can lead to scaling.

The effect of inlet pH on SO_2 removal in two different Shawnee lime scrubbers is shown in Figures 2.2.1-9 and 2.2.1-10. (Note that the effect of L/G discussed above can also be seen from these data.) Typical control points for a lime system are in the 8-9 range. Limestone systems, on the other hand, are normally operated at lower pH with typical control points in the 5-6 range. Scrubber inlet pH *versus* SO_2 removal are available for various magnesium ion concentrations (Figure 2.2.1-11) and various L/G ratios (Figure 2.2.1-12) for the Shawnee TCA scrubber. Both figures show increased SO_2 removals with increasing pH over the pH 5-6 range.

Inlet SO_2 concentration - For fixed design and operating conditions, greater SO_2 removal efficiencies are achieved at lower inlet SO_2 concentrations. This occurs because the actual quantity of SO_2 removed per volume of gas is

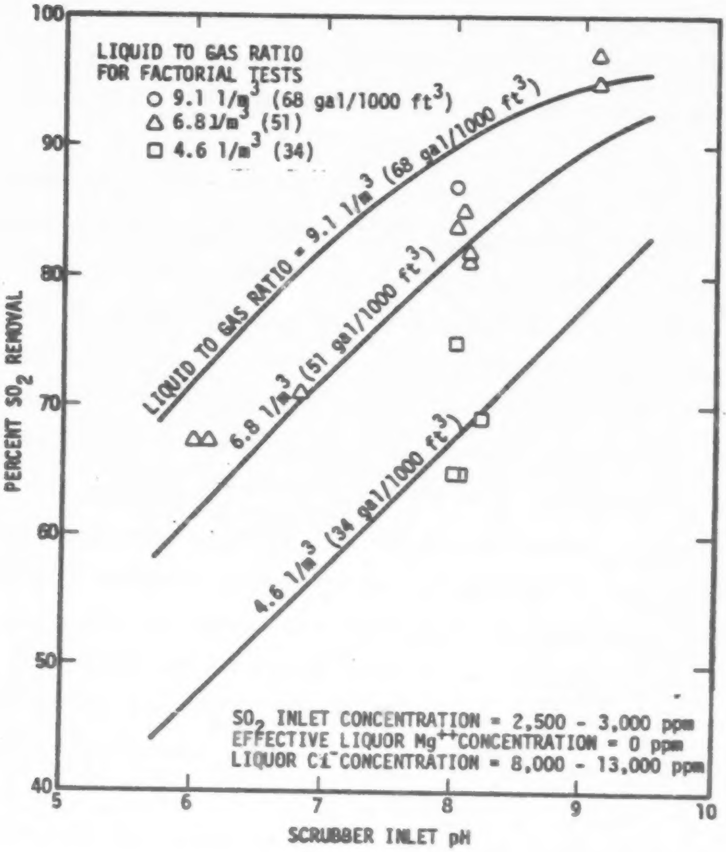


Figure 2.2.1-9. Scrubber inlet pH and liquid-to-gas ratio *versus* predicted and measured SO_2 removal, spray tower with lime, Shawnee plant.[45]

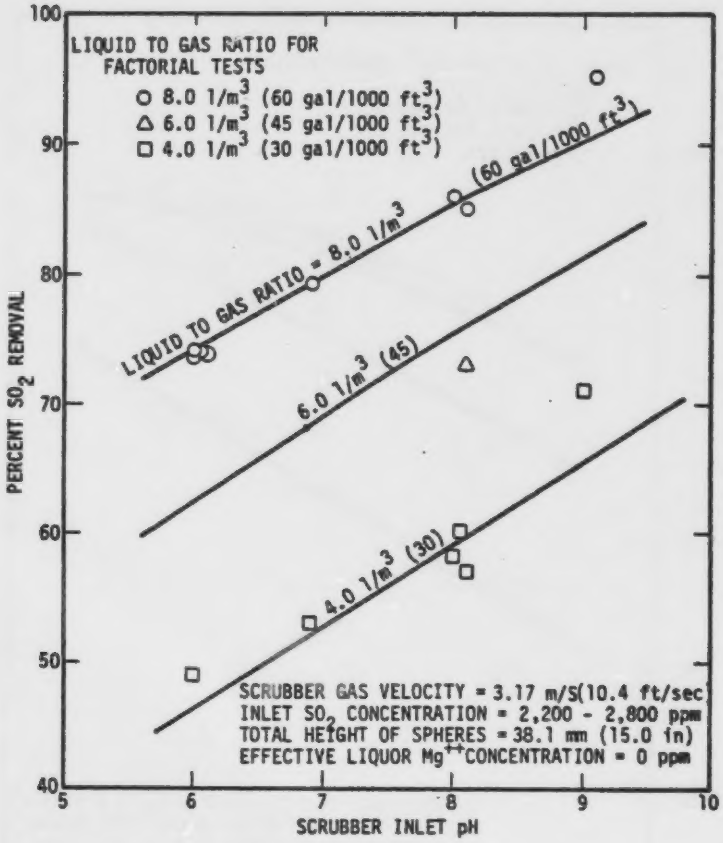


Figure 2.2.1-10. Scrubber inlet pH and liquid-to-gas ratio *versus* predicted and measured SO_2 removal, TCA with lime, Shawnee plant.[46]

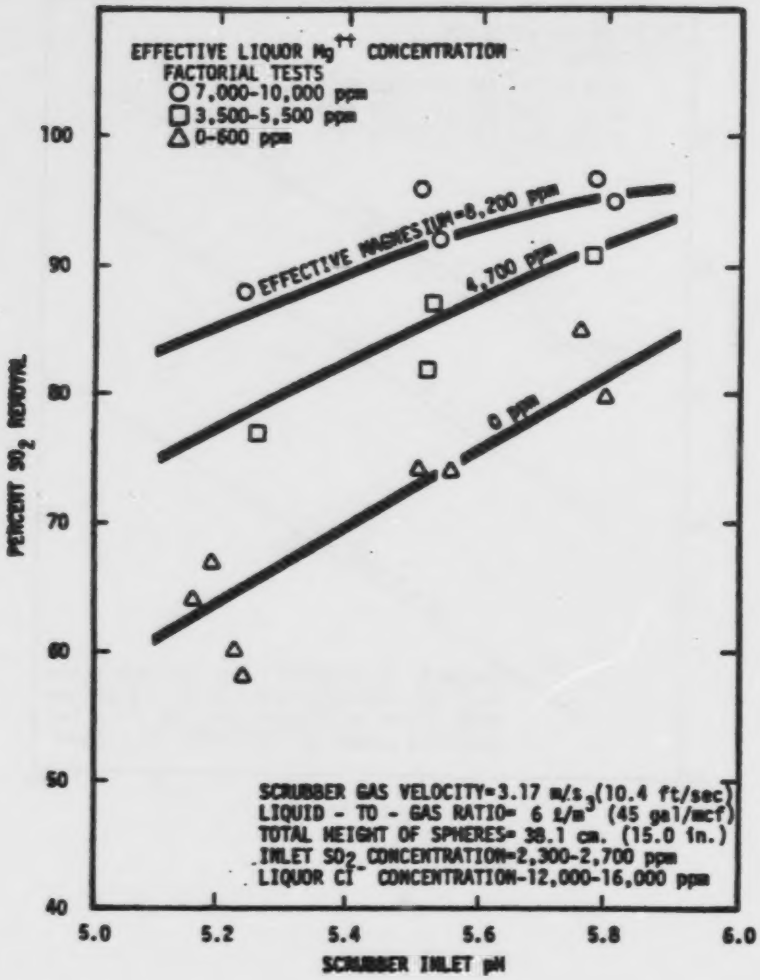


Figure 2.2.1-11. Scrubber inlet pH *versus* percent SO_2 removal at various magnesium ion concentrations TCA with limestone - Shawnee plant.[47]

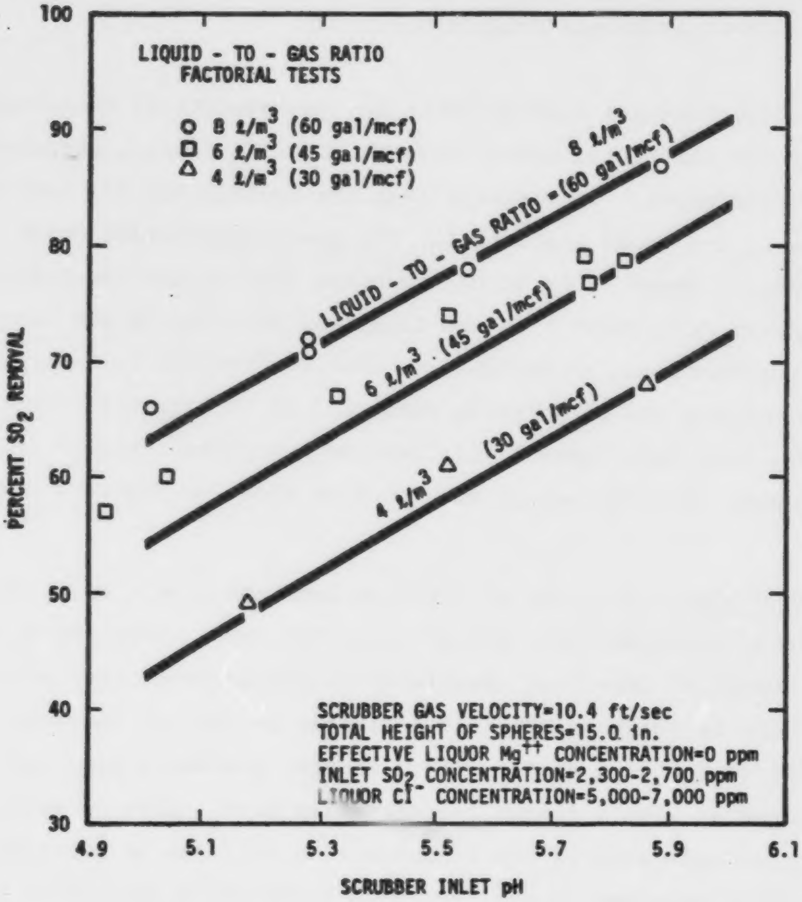


Figure 2.2.1-12. Scrubber inlet pH *versus* SO_2 removal for three L/G ratios -TCA unit with limestone - Shawnee plant.[48]

less, thereby reducing the load on the absorbent liquor. This is evidenced by several sets of limestone scrubbing data from Shawnee obtained with the TCA and spray tower scrubbers (Figure 2.2.1-13).

It was at one time thought that as inlet SO_2 concentrations decreased, the driving force for absorption would decrease and high removal efficiencies would be more difficult. It was felt that the equilibrium SO_2 back pressure over the bulk liquor would approach the SO_2 gas concentration which would limit absorption. However, in actual practice this is not the case since SO_2 back pressure over fresh lime and limestone slurries is very low (about 1 ppm), and scrubbers can be designed so that the low SO_2 concentration scrubbed gas exiting the scrubber is contacted by the freshest slurry.[51] For these reasons, very high removal efficiencies have been achieved from flue gas that averaged only 200 ppm at the inlet as shown in Figure 2.2.1-14.

Reaction tank design - The size of reaction tank can have a major impact on the amount of limestone that dissolves in the tank. This, in turn, affects the amount of dissolved alkalinity in the scrubber feed stream. A given increase in volume will yield a greater percentage increase in limestone utilization in a system with a shorter residence time tank than it will in a system with a longer reaction time tank. Also of major importance is the particle size of the limestone. A decrease in particle size will result in an increase in the limestone dissolution rate.[52, 53]

The operating conditions of the scrubbing system can also have a significant impact on the effects of hold tank size changes. If the scrubber inlet pH is lowered, the limestone dissolution rate will be enhanced and shorter residence time tanks can be employed. At Shawnee, limestone utilization was not affected by hold tank residence time changes when the scrubber inlet pH was maintained below 5.8. At higher pH's, a 6-minute residence time tank yielded significantly lower utilizations than did 12- and 20-minute residence time tanks.[54]

Hold tank configuration has also been shown to have an effect on both limestone utilization and SO_2 removal. Borgwardt's results indicate that plug

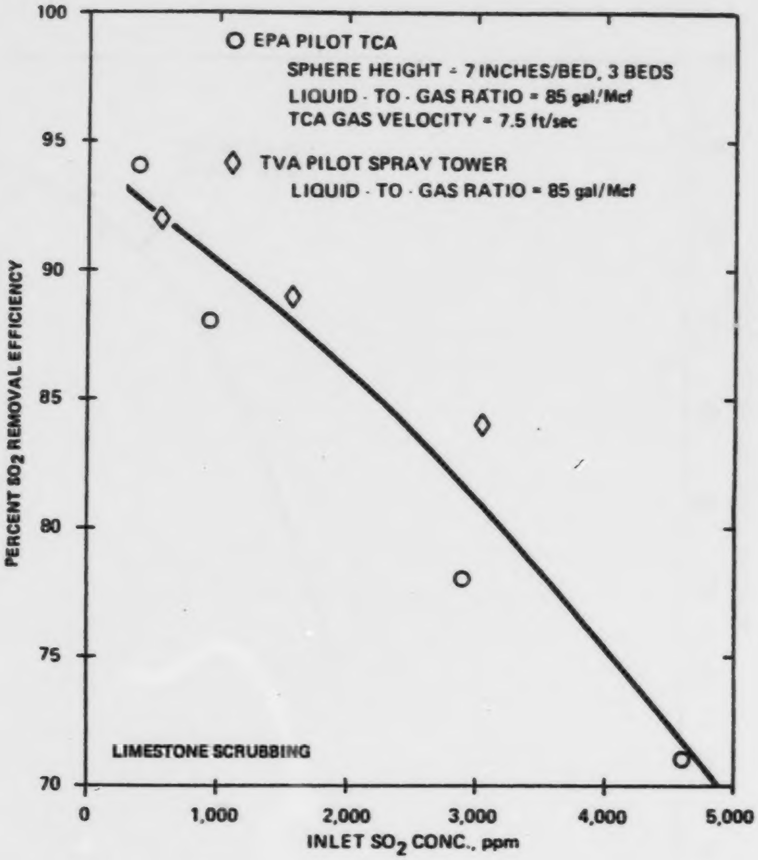


Figure 2.2.1-13. Effect of inlet SO_2 concentration on SO_2 removal efficiency for fixed design and operating conditions.[49]

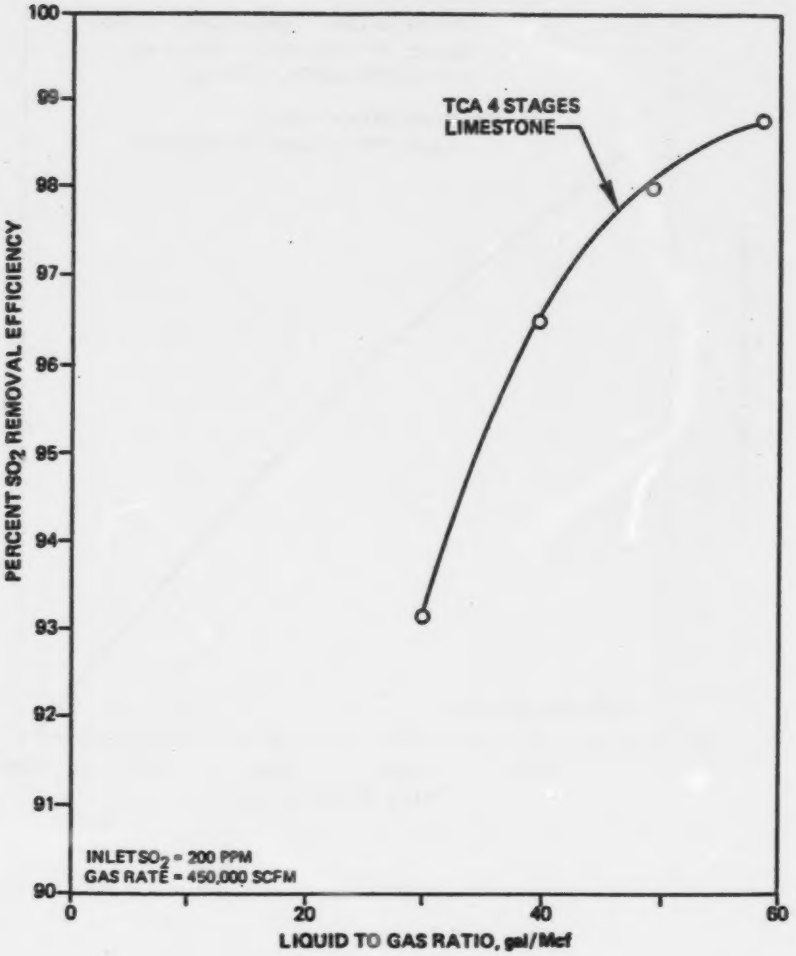


Figure 2.2.1-14. Effect of liquid-to-gas ratio on SO_2 removal efficiency with low sulfur coal at the Mohave power station.[50]

flow reaction tank designs can yield significant improvements in limestone utilization and SO_2 removal.[55] (A plug flow design is one that allows the reacting stream to flow through the reactor such that there is no backmixing. A plug flow situation can begin to be approximated by a number of mixed tanks in series.) At Shawnee, at a constant limestone addition rate, the SO_2 removal increased from 70% to 79% by changing the reaction tank from a single stirred tank to three tanks of equal total volume in series. This plug flow effect apparently drives the additive dissolution reaction further toward completion and makes more liquid phase alkalinity available for reaction with absorbed SO_2.

Effects of soluble species - One of the more important factors in determining liquid phase alkalinity is the distribution of the dissolved ions. Some of the species occurring in lime/limestone systems are soluble and exit the system only as dissolved solids. Important examples are Na^+, Mg^{++}, and $C\ell^-$. These soluble ions can enter the system as NaO or MgO in the ash or $HC\ell$ in the flue gas. They can also enter or be added to the system with the sorbent.

The soluble species determine to a large extent the levels of total calcium and total sulfur which remain in solution since the Ca^{++}, $SO_3^=$, and $SO_4^=$ species are constrained by solubility relationships. If Mg^{++} and Na^+ ion concentrations exceed the $C\ell^-$ ion concentration (on a normality basis), there will be more soluble cations in solution than anions. To satisfy electrical neutrality requirements in this situation, larger amounts of basic anions such as $SO_3^=$, and $SO_4^=$ will be required to remain in solution. Because the amounts of Ca^{++}, $SO_3^=$, and $SO_4^=$ in solution are controlled by solubility constraints, solutions which are high in Mg^{++} and Na^+ will generally contain large amounts of sulfur and relatively little calcium. On the other hand, if $C\ell^-$ concentrations exceed Na^+ and Mg^{++} concentrations, there will tend to be a higher relative concentration of Ca^{++} ions in solution.

Addition of relatively small amounts of soluble magnesium species (less than 1 percent by weight) has been shown to be beneficial to SO_2 removal efficiency. Since the magnesium compounds are much more soluble than the calcium-based reactants, the SO_2 can react rapidly with the active liquid-phase magnesium species, thus making the limestone dissolution rate much less of a determining factor in the SO_2 removal efficiency. This is illustrated in Figure 2.2.1-15, and is substantiated by data from both pilot and full-scale systems. Magnesium addition was tested at two different inlet SO_2 concentrations (medium and high) on Combustion Engineering's 0.56 m^3/s (1200 cfm) pilot plant using lime as the scrubbing agent. These results are presented in Figure 2.2.1-16. Some data from a small limestone system (Shawnee TCA) are shown in Figure 2.2.1-17.

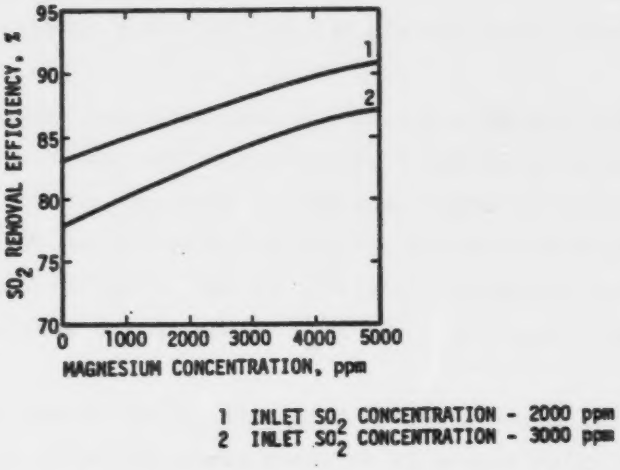


Figure 2.2.1-17. Effect of magnesium on SO_2 removal efficiency.[58]

Test results from two full-scale systems have also been reported.[59] At Paddy's Run, magnesium addition increased SO_2 removal efficiency from 83 percent (average) to 99.7-99.9 percent; the inlet SO_2 concentrations ranged from 2150-2230 ppm. At Phillips Station, which is also a lime system, magnesium addition resulted in SO_2 removal increasing from 50 percent to 83 percent. The source of magnesium was lime containing up to 10 percent Mg. It should be pointed out that magnesium must be continually added to the system to make up for that lost in the adherent sludge liquor and that tied up with chloride ions present in the system.

Number of contact stages - Another way to increase SO_2 removal is to increase the number of contact stages in the FGD system. This provides a means to contact the SO_2-laden gas with a fresh, unsaturated scrubbing slurry after

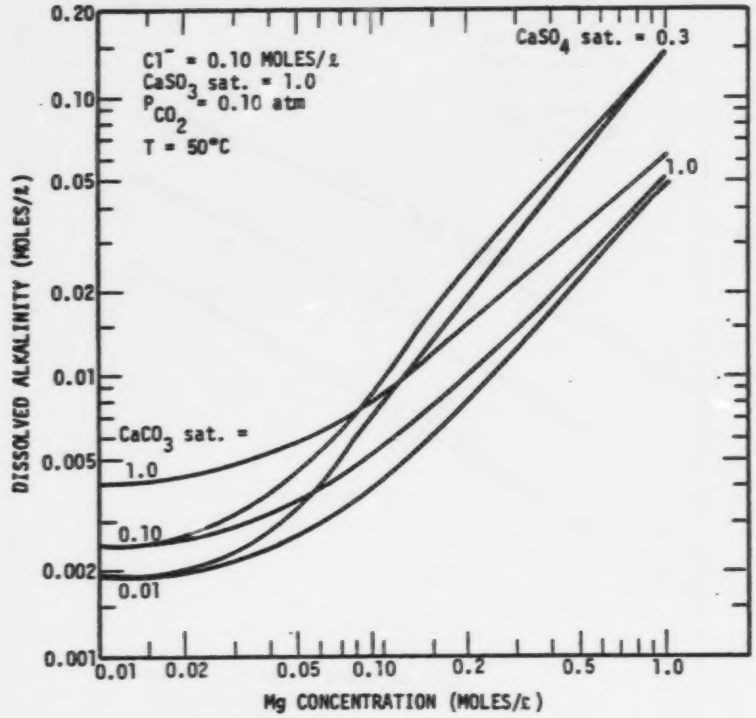


Figure 2.2.1-15. Dissolved alkalinity generated by addition of MgO.[56]

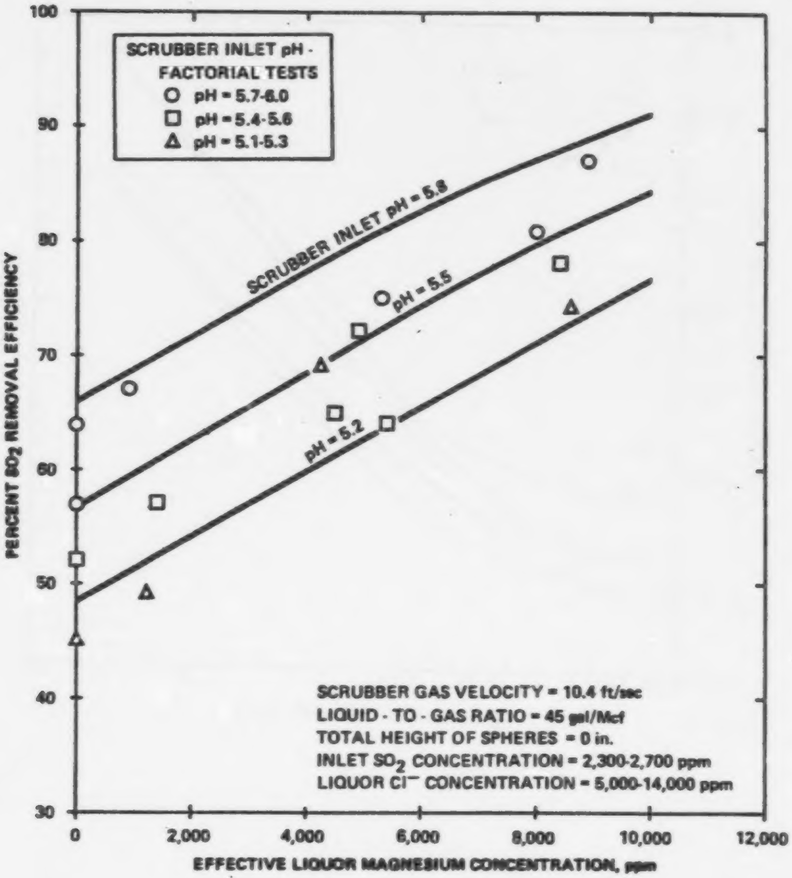


Figure 2.2.1-16. Effect of Magnesium on SO_2 removal efficiency - TCA (no spheres) with limestone.[57]

the first stage of SO_2 removal is complete. This effect tends to level off as the number of additional stages is increased, as shown by some Mohave lime scrubbing data obtained on the horizontal module shown in Figure 2.2.1-18. It was also reported that a similar effect was measured in the TCA scrubber at the Shawnee test facility when two different scrubbers were operated in series.[61] The results, said to be generally true for both lime and limestone systems, are shown in Figure 2.2.1-19.

Ash removal - An important design consideration for coal-fired system applications is whether or not to remove particulates upstream of the scrubber. The current trend within the utility industry is installation of dry particulate collection equipment upstream of the FGD system. With low-efficiency precipitators (90-95 percent removal) or mechanical collectors, the particulate emission control costs may be considerably reduced, but the scrubber design must be capable of removing residual particulate.

Some scrubber types (venturi or mobile-bed) can effectively control both particulate and SO_2, in some cases with a two-stage arrangement. Although the capital cost may be minimized and the ash may contribute alkalinity to the system, there are several significant disadvantages associated with removing particulates in the FGD scrubber.[63]

1. The extent to which the sludge can be dewatered by addition of dry fly ash is reduced. The importance of this factor depends on the sludge disposal method to be used.
2. There is a general consensus that ash causes erosion in the scrubber; on the other hand, some degree of erosivity may be desirable to keep the internal surfaces free of scale and deposits.
3. To avoid incidences of exceeding particulate emission regulations, by-passing the scrubber would be questionable. As the reliability of lime/limestone systems continues to increase, this factor diminishes in importance.
4. Fly ash cannot be marketed unless collected dry upstream of the scrubber.
5. Particulate scrubbing results in an increased pressure drop which in turn increases power consumption. This is felt not only as an operating cost item, but also as an increased requirement for

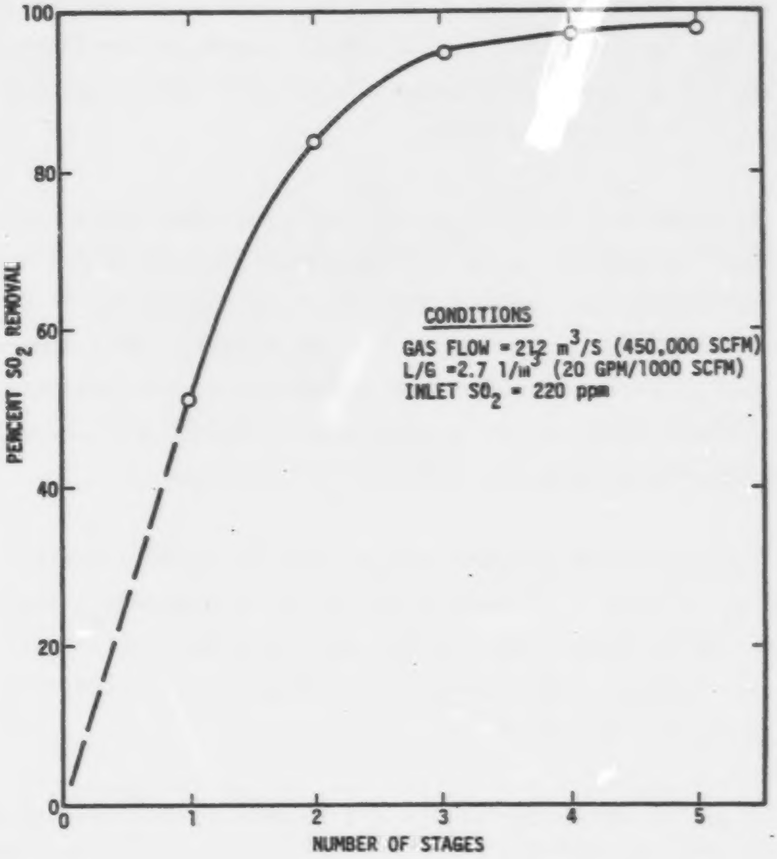


Figure 2.2.1-18. 170 MW horizontal SO_2 removal *versus* number of stages, Mohave plant.[60]

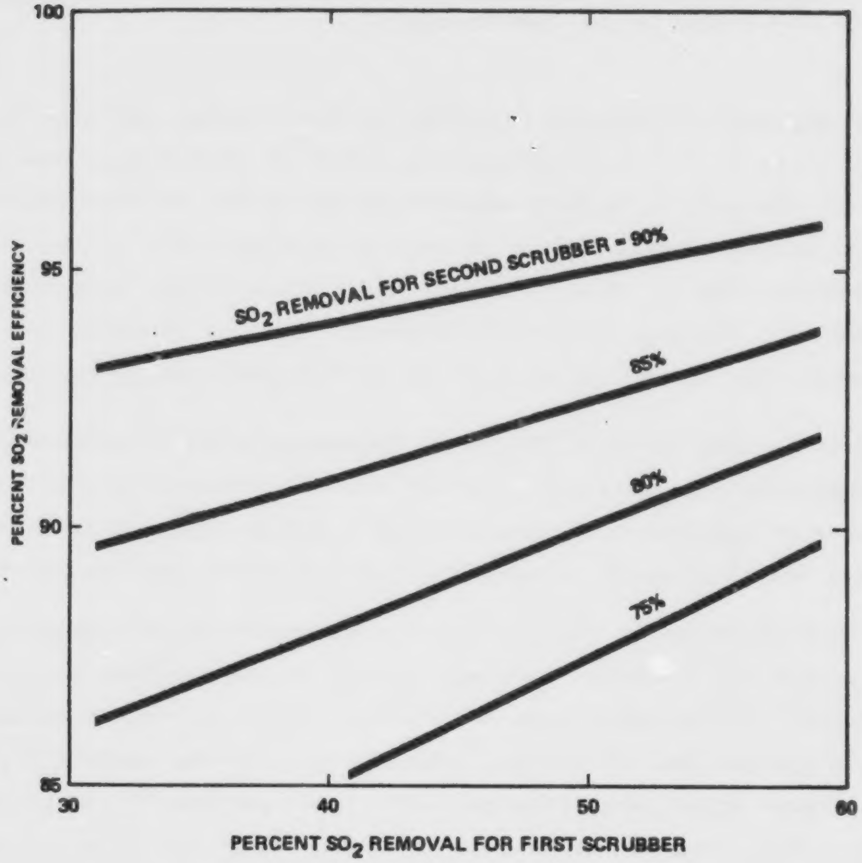


Figure 2.2.1-19. SO_2 absorption efficiency for two scrubbers in series.[62]

overall boiler capacity. On the other hand, burning low-sulfur coal requires a relatively high power requirement for the precipitator anyway, thus making combined particulate/SO_2 scrubbing more suitable for western coals.

Oxidation and scale prevention - Scaling in the scrubber can be a function of several operations in the scrubbing loop. Low L/G ratios may cause scaling by allowing the calcium sulfate supersaturation in the scrubber liquor to exceed the critical level. This can lead to precipitation of calcium sulfate on scrubber internals. This phenomenon would occur if the hold tank were undersized, not allowing sufficient residence time for precipitation and desupersaturation of calcium sulfite and sulfate and dissolution of absorbent.

The rate of scaling is sensitive to the supersaturation of calcium sulfate. If supersaturation is unchecked, calcium sulfate dihydrate crystals start to crystallize on equipment surfaces, forming a scale. Calcium sulfate supersaturation levels of up to 1.3 can be tolerated before scaling occurs.

Calcium sulfate formation results from the oxidation of calcium sulfite. For a given system the oxidation rate will depend on the relative concentrations of O_2 to SO_2. For example, some systems applied to low sulfur western coals have shown greater than 90 percent oxidation whereas the eastern coals exhibit lower rates which range from near zero to 40 percent.[64] Consequently, in predicting oxidation, it has been observed that this rate is a strong function of the oxygen content of the flue gas. This observation must be considered when designing for systems with relatively high O_2/SO_2 ratios, such as are encountered in industrial boilers, as well as those where this ratio might vary. A variation in O_2/SO_2 ratio would be observed in systems which encounter large changes in excess air in the boiler operation. Therefore, a system which must accommodate significant changes in flue gas concentration should be designed for longer residence times in the hold tank and higher L/G ratios.

Water balance - Another important design consideration is the water balance within the system and its role in the overall plant water management system. The following factors impact the design of the water system:

- Evaporation loss in the scrubber
- Adherent liquor discharged with the sludge
- Pump seals
- Mist eliminator wash
- Pond considerations

Water loss via evaporation in the scrubber, generally 0.4-0.5 tons/ton of coal fired, allows for fresh water addition to the system. Fresh water input is generally used for mist eliminator washing, slurry preparation, and sometimes pump seals. During periods of low boiler load, however, evaporation loss drops considerably. Fresh water makeup requirements are not reduced to the same degree, thereby creating a problem in maintaining the proper water balance.

Adherent liquor discharged with the sludge constitutes the other main water loss from the scrubbing system. The quantity varies considerably from site to site depending on the type of dewatering used, and the amount and type of solids discharged (*i.e.*, sulfur-containing species and ash). Water loss can range from 100 lb per ton of coal burned for a case using highly efficient dewatering methods to as much as ½ ton or more of water per ton of coal.[65]

To avoid erosion of pump seals in slurry systems, relatively clean water is needed. This requirement can be met either by fresh water or clarified recycle water. Another fresh water requirement is for mist eliminator wash, although in some cases a mix of recycled clarified overflow with fresh water has been successfully used.

Several features of pond design and operation are also important to the overall water balance. For example, rainfall surface runoff, evaporation loss and leaching make design of completely closed loop operations site specific.

Gas Reheat - In general, industrial boiler FGD processes do not use stack reheat systems whereas utitity FGD applications do.[66,67] This discussion briefly summarizes the reasons for applying reheat, and addresses potential reheat mehtods.

Flue gases are normally discharged from a utility boiler at a temperature of greater than 120°C (250°F). This temperature is selected to remain above the H_2SO_4 dew point to reduce corrosion and permit use of less expensive grades of material to be used for fans, ducts, and stacks. Industrial boilers generally have less efficient heat recovery systems and, consequently discharge flue gas at a temperature of 175-200°C (350-400°F).[68] When a wet scrubber is inserted between the boiler outlet and stack, the flue gas exiting the scrubber is saturated with water and cooled to the adiobatic saturation temperature of about 50°C (125°F). Discharging the cool wet gas directly to the atmosphere can result in the following problems:

- Condensation of water vapor and sulfur oxides, which can result in the acidic water corrosion of downstream ducts, fans, and stack lining.
- Impaired plume rise and, hence, poorer dispersion of residual pollutants for a given stack height.
- Deposition of scrubber residue on downstream fan blades, resulting in imbalance.
- A visible plume as water vapor condenses.
- Stack rain, or mist droplets, that can settle around the power station.

To prevent these undesirable effects of wet scrubbers, the exit gas can be reheated to a higher temperature before being discharged.

Several approaches have been developed to reheat scrubber gases The basic differences among these approaches are in the energy sources and the method of transferring energy to the gas. Reheat methods currently in use include:

- Direct inline reheat - using steam or hot water exchangers
- Direct combustion reheat - using gas or oil in inline burners or external combustion chambers
- Indirect hot air reheat - using steam to heat air to mix with the scrubbed gas
- Bypass reheat - bypassing a portion of the untreated hot flue gas to mix with the scrubbed gas

Although the majority of utility FGD installations use reheat, there are several installations which operate under "wet-stack" conditions as do the majority of industrial FGD systems. Operating without reheat minimizes the energy consumption of an FGD system.[69] Current reheat practice is to provide about 30°C (50°F) of heat to overcome problems associated with discharging wet gas to the atmosphere. This practice results in an energy penalty of about 2 percent of the net heat input to the boiler. Although reheat has the potential for providing significant benefits, it should not be considered as a necessity, but as one of a number of approaches for consideration in optimizing sulfur dioxide absorption systems.

Sludge dewatering and disposal - The sludge dewatering process step is used to concentrate the solids for ease of handling and disposal and to lower transportation costs. The clear liquor is usually recycled back to the process for reuse. The methods currently used in lime/limestone sludge dewatering on commercial-sized units are thickening and vacuum filtration. Centrifugation was also tested, but filtration has been the selected method for full-scale applications in most cases. In addition to these methods, interim ponding has been used as a dewatering procedure.[70]

Clarifiers or thickeners are commonly used as a primary dewatering device for sludges containing low solids content (10-15 percent solids). They can typically achieve a solids content of 30-35 percent. Depending on the dewatering properties of the sludge and the ultimate disposal plans, further steps may be taken to increase the solids content using vacuum filters, which achieve 50-75 percent solids, depending on the individual system. Because of the physical nature of sulfate crystals compared to sulfite, dewatering is generally improved by a higher sulfate/sulfite ratio. A sludge from a forced oxidized system, which contains mostly calcium sulfate can achieve about 85 percent solids using either centrifugation or filtration.

The dewatering method selected is heavily influenced by the type of disposal to be used. While the basic options are wet (i.e., ponding) or dry (i.e., landfill), the composition of the sludge and the availability of dry fly ash to supplement dewatering are also important factors.

Erosion and Corrosion - Early FGD installations experienced many shutdowns due to corrosion and erosion. While significant advances have been made in solving materials problems, some material failures are still reported and studies are still underway to investigate the metallurgical aspects of FGD.

In recent reports, several guidelines for selection of materials for FGD systems were laid out based on experience to date.[71,72] The preferred material for all wet surfaces, including the FGD module itself as well as inlet and outlet ducts, is 316L stainless steel. Its superior performance with respect to resistance to corrosion, erosion and scaling may be due to its molybdenum content. In comparative metallurgical tests, 316L has proven better than carbon steel, low alloy steels, type 304, and type 304L. There are some systems, however, that incorporate such alloys as Hastelloy C-276, Hastelloy G, Inconel 625, Incoloy 825, 317L stainless steel, 904L stainless steel, and Jessop JS700. These materials have been used in wet/dry high temperature, high chloride environments, such as in presaturators.

It is important that pumps, lines, and recycle tanks in wet slurry systems be protected from erosion. Materials successfully used in these key elements include synthetic and natural rubber coatings.

Lining materials for the absorbers, exhaust ducts and stacks have also been tested. Although successful applications have been reported, some liner failures have been due to improper application, instability at high temperatures, inconvenience of repair, and cost-related factors. Materials successfully used to date in this type of application include resins, ceramics, polyesters, polyvinyls, polyurethanes, carboline and Gunite.[73]

2) Fuel variations - The design and operation of lime/limestone scrubbing systems are affected by variations in the fuel characteristics listed below:

- Sulfur content
- Chloride content
- Ash alkalinity
- HHV

The effects of these variables on the design of lime/limestone systems are much like the effects on other wet FGD systems. The following discussion describes the importance of these fuel variables on the design of lime/limestone wet scrubbing systems.

Sulfur content - The sulfur content of the fuel determines the removal rate of SO_2 which is necessary to comply with a given environmental regulation. The design of virtually every piece of equipment is affected when attempting to achieve a set emission rate with fuels of varying sulfur content. Increasing the fuel sulfur content affects the design of the absorber in that higher liquor circulation rates are required for higher sulfur fuels which results in bigger pumps and piping, and an increased scrubber pressure drop with increased system energy requirements. In addition, larger alkali storage, preparation, and feed equipment is required to handle the increased alkali necessary to react with the larger amounts of sulfur. Finally, larger solids separation and waste sludge disposal equipment will be needed to handle the increased load. In summary, FGD systems designed for high sulfur fuels will be more complex and costly than systems designed for low sulfur fuels.

Chloride content - During combustion the chlorine content of the fuel forms gaseous HCl which will be readily absorbed in aqueous FGD systems. Chlorine is important both for corrosion reasons, and because it will react with the alkaline sorbent thus limiting the SO_2 removal ability of the system. The presence of chlorides in the scrubber liquors provides the potential for stress corrosion which can result in the use of high-alloy equipment wherever protective coatings are not applicable.

Dissolved chlorides will also react with active alkalis to form inactive chloride salts, and consequently a high chloride coal will require a significantly larger liquor circulation rate than will a coal with a lower chlorine content to achieve the same amount of SO_2 removal. The use of alkali additives, such as MgO, may be used to increase the dissolved alkalinity which will permit SO_2 removal without an increased liquor circulation rate. Alternatively, chlorine can be removed ahead of the SO_2 scrubber in a relatively small prescrubber.

Ash alkalinity - Another fuel variation that can affect the design for wet FGD systems is the ash alkalinity. As previously discussed, a highly alkaline ash can significantly decrease the quantity of sorbent required for SO_2 removal. Commercial scale utility systems are in operation today which are based on a combined sorbent-alkaline ash design. The ash alkalinity will influence the design consideration of whether or not to remove particulate material ahead of the scrubber.

HHV - A boiler using a low heating value fuel must fire at a higher burn rate to generate the same amount of power as for a high heating value fuel. The low heating value fuel will produce a larger volume of flue gas per unit of power than the high heating value fuel which will result in the FGD system being of a larger size to accomodate the greater gas volumes. In addition, if both fuels had the same sulfur content, the low heating value fuel would result in greater SO_2 emissions per unit of power.

3) Ambient variations - FGD system operations are essentially independent of ambient variations. However, as with all wet systems, extreme cold can adversely affect the operation of a lime/limestone FGD process. Problems with line freezing and subsequent unit shutdowns were experienced at several operating facilities in the East during the harsh winter of 1977. These problems can be solved by heat tracing lines or by locating the FGD system inside buildings. Also, climatological factors may effect visible plume formation and should be taken into consideration on reheat decisions.

E. Retrofits--Existing boiler operating parameters such as flue gas temperature and oxygen content will definitely affect the design of a lime/limestone system. There are no specific conditions other than equipment space requirements, however, which would actually prevent application of a lime/limestone flue gas cleaning process.[73]

Major process equipment for a lime/limestone system includes the scrubber where contact between the flue gas and slurry is promoted, a mist eliminator for removal of entrained liquid from the SO_2-lean gas, a reaction tank where additional lime/limestone dissolution and solid product precipitation occurs, and a solids separation device to reduce the amount of liquid waste leaving the process. Flue gas reheat may also be used on some installations. In addition to the equipment listed above, several other items are needed. Storage bins for the alkaline additive are needed. Slurry tanks and slurry pumps are also required. A waste sludge handling area is required for both the lime and limestone systems. Depending on whether on-site solids disposal is planned, equipment in this area could include a clarifier or thickener. vacuum filters or sludge fixation equipment. Space may also be needed for sludge storage.

Space required for the scrubbing section of this process is the major concern for retrofit since this equipment must be placed adjacent to the powerhouse and stack. Process equipment outside of the scrubber area is of less concern to the retrofit problem since it can be located on the peripheral areas of the plant. By employing good design practices and placing the hold tanks under scrubbers, space requirements for the scrubbing section can be minimized. However, for the large industrial boiler application considered in this evaluation, 200 X 10^6 Btu/hr, a scrubber diameter of only 3M (10 ft) would be sufficient to provide a gas velocity of 3M/S (10 ft/sec). This relatively small scrubber size does not appear to be large enough to limit its applicability.

2.2.1.2 System Performance--

A. Emission reductions--A discussion of design and operating parameters for lime and limestone FGD systems was presented in Section 2.2.1.1.D. Data for that discussion came primarily from EPA's 10 MW test facility located at TVA's Shawnee steam plant. In this section, results of process testing recently completed on the Bahco Process installed at Rickenbacker Air Force Base (RAFB) will be presented to illustrate the performance of a lime/limestone FGD system applied to industrial boilers.

In the Bahco Process, flue gas is passed through a mechanical collector to remove coarse particulate matter before entering the booster fan. The booster fan forces flue gas into the first stage of the scrubber where it is vigorously mixed with scrubbing slurry in an inverted venturi. In this stage, flue gas is cooled to its adiabatic saturation temperature and SO_2 and particulates are scrubbed from the gas. This partially scrubbed gas rises to the second stage where it is contacted with slurry containing fresh lime to complete the required SO_2 and particulate removal. Gas from the second stage enters a cyclonic mist eliminator where entrained slurry droplets are separated from the gas by centrifugal force.

Pebble lime from a storage silo is slaked and added directly to the slurry in the lime dissolving tank. The resulting fresh lime mixture is pumped to the second stage (upper) venturi to treat the flue gas stream. The slurry flows by gravity from the second stage to the first stage where it contacts hot flue gas entering the scrubber. This countercurrent flow arrangement results in high SO_2 removal and efficient reagent usage.

Spent slurry flows by gravity from the first stage of the scrubber to the dissolving tank. Part of the spent stream leaving this stage is diverted to the thickener where the slurry is concentrated to 35 to 40% solids. Overflow from the thickener returns to the dissolving tank and the underflow is pumped to a Hypalon-lined sludge pond near the Heat Plant. The slurry and gas streams described above are illustrated in Figure 2.2.1-20.

The Bahco system was designed to treat up to 108,000 acfm of flue gas generated at the peak winter load of approximately 200 X 10^6 Btu/hr. The system, which must operate over a relatively narrow range of gas flow, 35,000 to 50,000 scfm, has an essentially unlimited turndown capability for handling flue gas by mixing air with the flue gas at low boiler loads. This allows the system to handle seasonal load variations from 20 to 200 X 10^6 Btu/hr, SO_2 concentrations from 200 to 2000 ppm, and partiuclate loadings up to 2 gr/scfd. In addition, the scrubbing system has coped with 100 percent load increases occurring in as little as an hour's time. Table 2.2.1- 9 shows scrubber operating conditions during average load conditions for the RAFB installation.[74]

TABLE 2.2.1- 9. TYPICAL BAHCO OPERATING CONDITIONS[76]

Flue Gas	64,000 acfm @ 380°F (37,500 scfm)
SO_2 Concentration	1390 ppm
1st Stage ΔP	10 in W.C.
2nd Stage ΔP	8 in W.C.
Lime-SO_2 Stoichiometry	0.876
SO_2 Removal	87.6% (0.615 lb/10^6 Btu)
Lime Utilization	100%
Particulate Emission	0.16 lb/10^6 Btu

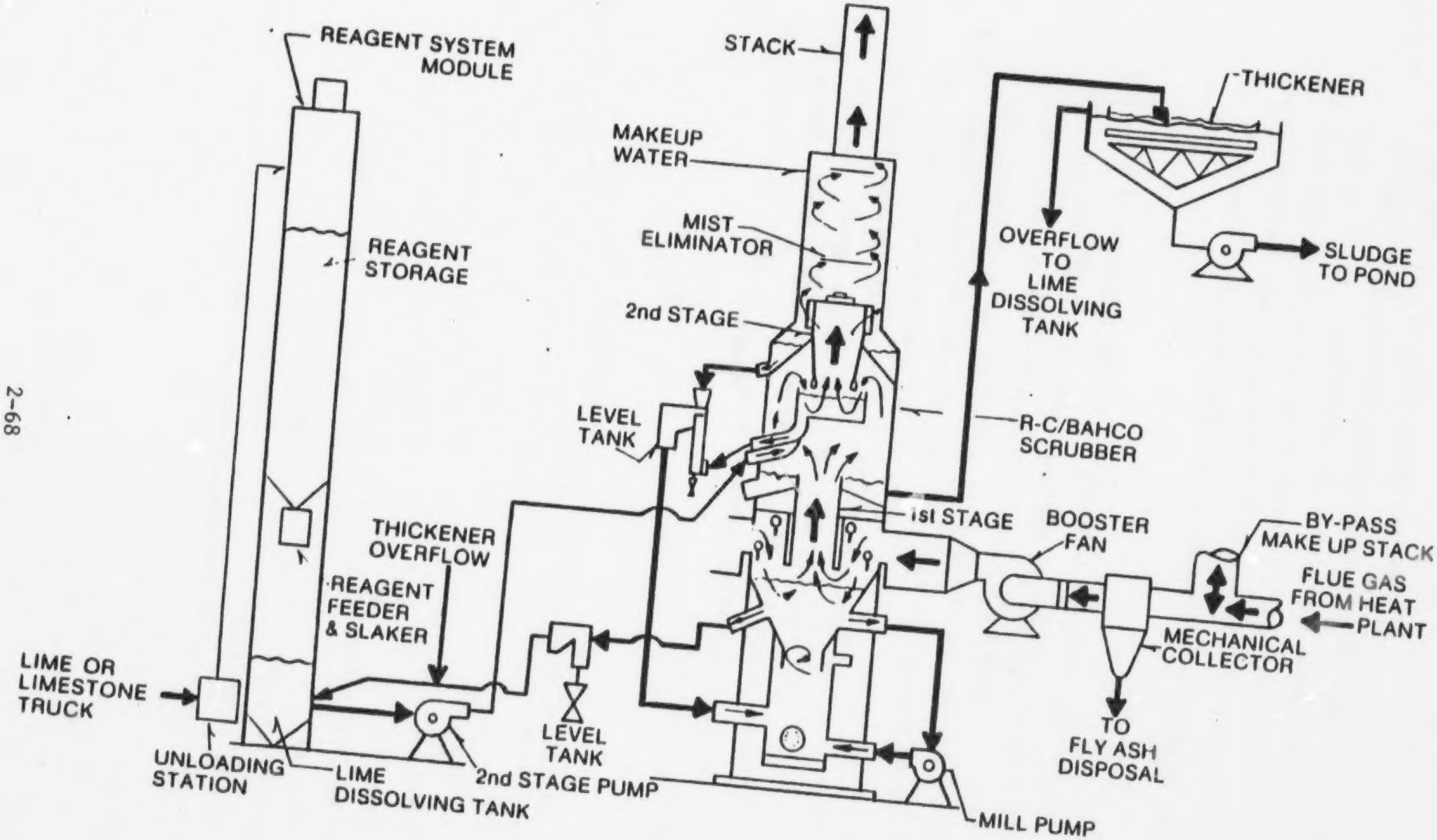


Figure 2.2.1-20 R-C/Bahco scrubber system flow diagram[75]

During the test period, the system operated satisfactorily at slurry solids from 2 to 25 percent by weight. Particulate emissions were above the required level of 69 ng/J (0.16 lb/10^6 Btu) when substantial quantities of soot were being formed in the heat plant. Sulfur dioxide removal efficiency was well above the required level, and levels in the 95 percent range were attainable. The following discussion summarizes the SO_2 removal results obtained when operating with lime and limestone, the particulate removal ability of the system, and the properties of the by-product sludge.

Figure 2.2.1-21 shows a nearly linear relationship exists between lime/SO_2 stoichiometry and SO_2 removal efficiency. Lime utilization approached 100 percent in the range of stoichiometry from 0.3 to 0.9 moles lime/mole SO_2 and dropped gradually to 90-95 percent as SO_2 removal approached 100 percent. Over the range of liquid circulation rates studied, 1500-2400 GPM, no effect on SO_2 removal was observed. Results of these lime tests indicate that SO_2 removal is controlled by the lime-SO_2 stoichiometry. The fact that SO_2 removal when using lime is not adversely affected by changes in operating variables illustrates that good gas-liquid contact was maintained over the entire operating range of the system.[77]

Results of process testing using limestone indicated that slurry circulation rate, as well as limestone-SO_2 stoichiometry, are important in determining the SO_2 removal efficiency. Figure 2.2.1-22 shows the positive effect on SO_2 removal and limestone utilization of increasing the slurry rate from 2000 to 2600 gpm. This figure also shows that limestone utilization varied from 75 to 90 percent at lower SO_2 removal levels but decreased significantly above 80 percent SO_2 removal. The scatter shown in the data on Figure 2.2.1-22 were reported to be within the uncertainty limits of the data, and project scope limitations precluded running additional verification tests to investigate this further.[79]

The Bahco system at RAFB was designed with extra fan capacity in order to help carry out the particulate test program. Venturi pressure drops

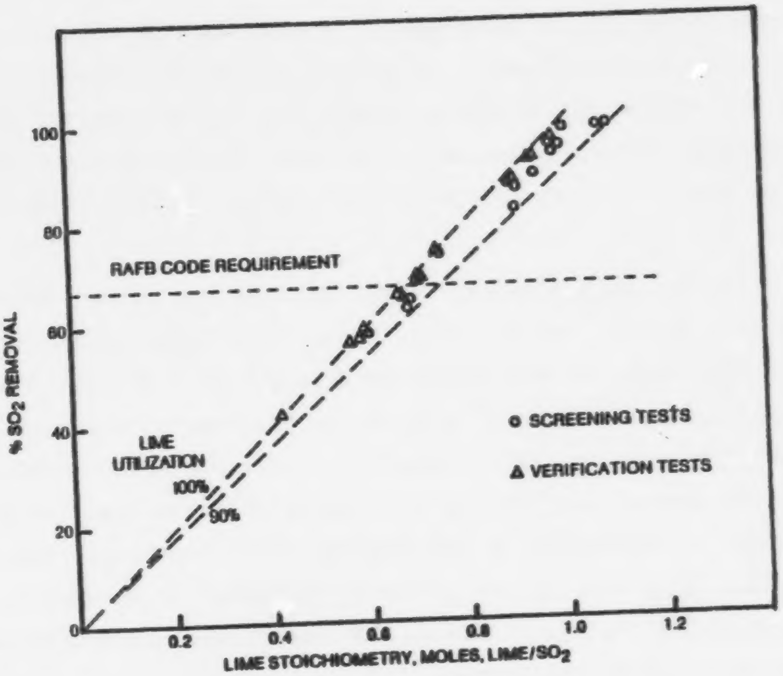


Figure 2.2.1-21. SO$_2$ removal efficiency as a function of lime stoichiometry.[78]

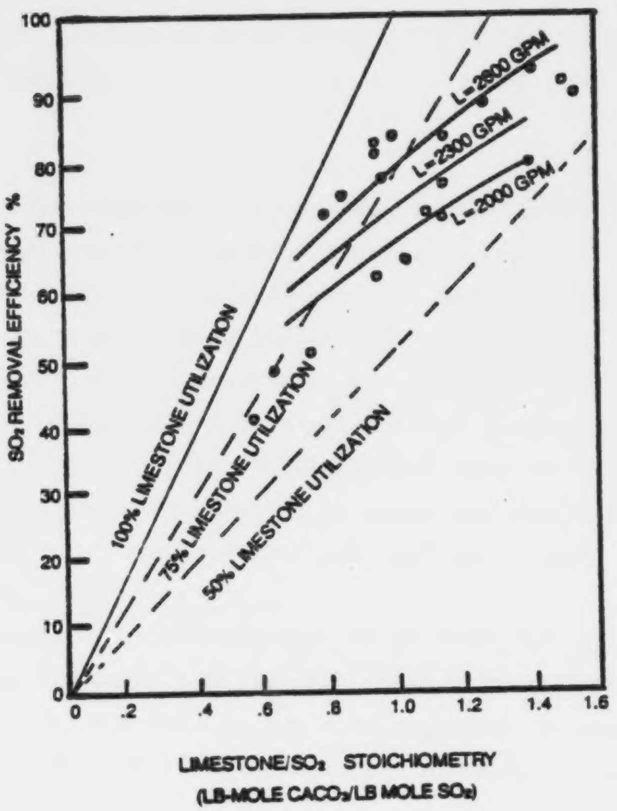


Figure 2.2.1-22. SO_2 removal efficiency as a function of limestone/SO_2 stoichiometry and slurry pumping rate.[80]

were increased to nearly double their design values of 7 in. W.C. for the particulate tests. Data from the particulate control tests are plotted in Figure 2.2.1-23 to show the effect of the combined pressure drop of the two venturies in the Bahco scrubber on particulate emissions. Below approximately 18 in. H_2O total pressure drop, particulate emissions increase rapidly.[82]

A series of scrubber sludge characterization tests were carried out to compare the lime and limestone sludges. In general, the limestone sludges settled much more rapidly than the lime sludges as shown in Figure 2.2.1-24. In both cases, the concentration of settled sludges were not affected by the addition of flocculant. However, the limestone sludges, due to their higher gypsum contents, produced a settled layer with a higher solids content (58 weight percent) than the lime sludges (43-45 weight percent).[83] Results of centrifugation and filtration tests as shown in Figure 2.2.1-25 also shows better dewatering results for the limestone sludges.[85]

B. Reliability--Reliability of lime/limestone FGD systems for industrial boiler applications is difficult to assess since there are only two installed systems and only one of those has been operational over a long period of time. That system is the Bahco system located at Rickenbacker Air Force Base.

Scrubber performance at the RAFB facility has generally been quite good except perhaps for the early stages of operation in which several start-up problems resulted in significant amounts of downtime. From the time period of November 1976 through December 1978, however, the Bahco system illustrated that an industrial boiler can operate with high reliability as it operated 95 percent or more of the time except for the months of January-March 1978. During those three months, system downtime was caused by a severe blizzard which resulted in the freeze-up of several lines. In February, 1978, the Governor of Ohio temporarily shelved pollution contorl regulations and the system was not operated until April at which time it was reported to perform with an operability index of 100 percent.[87]

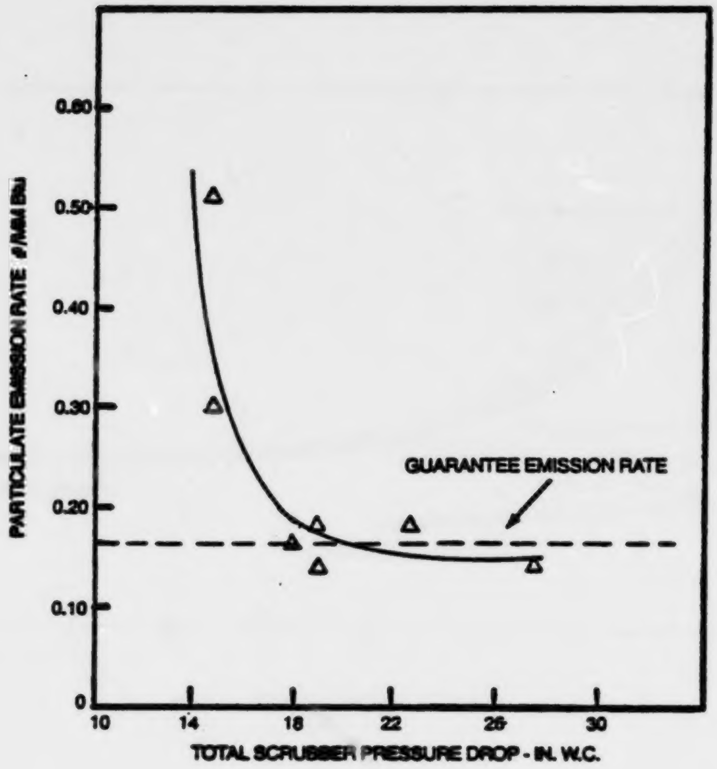


Figure 2.2.1-23. The effect of scrubber pressure drop on particulate emission rates.

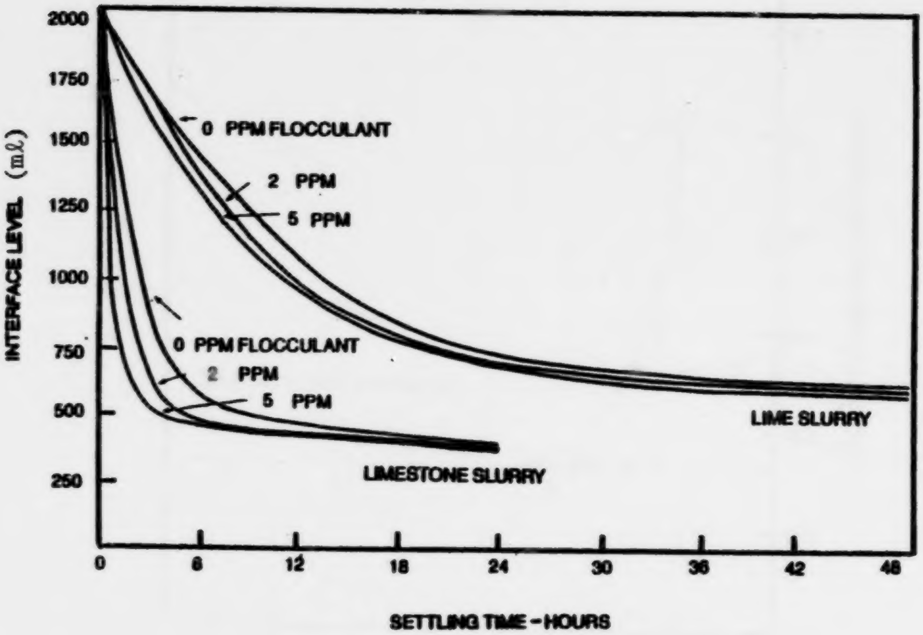


Figure 2.2.1-24. A comparison of lime and limestone slurry settling rates.[84]

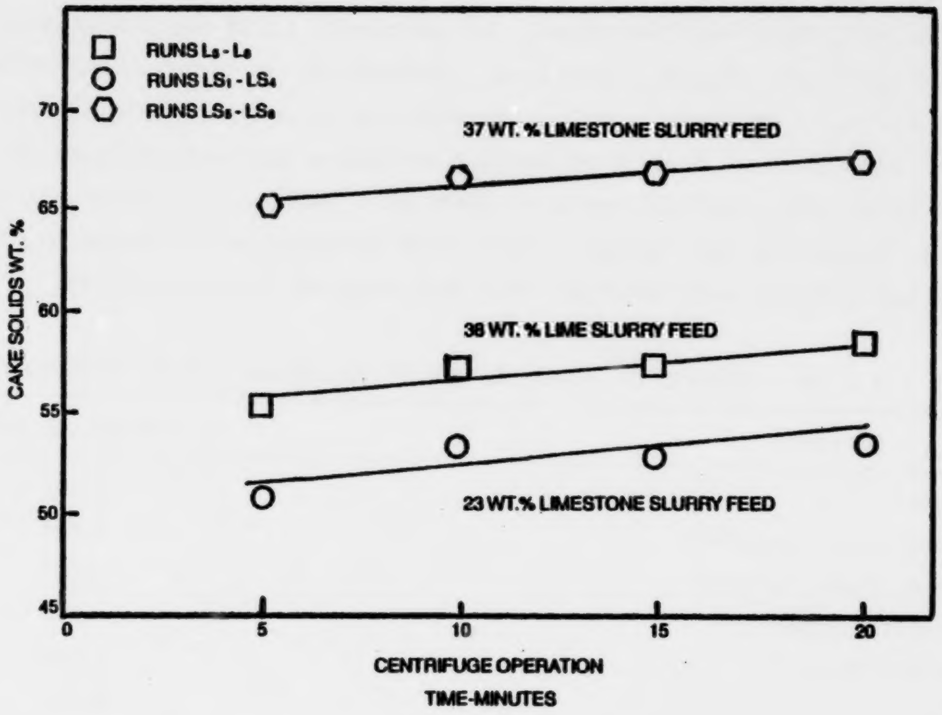


Figure 2.2.1-25. The effect of operating time and slurry feed concentration on centrifuge cake density.[86]

During the early stages of operation of this unit, several start-up problems were experienced that resulted in considerable amounts of downtime. During this period which included 11,024 hours, there were 4,830 hours of downtime. Most of this downtime, 2,766 hours, resulted from booster fan problems which have been rectified. An additional 1,088 hours were lost due to other auxiliary equipment problems. However, of the total of 3,854 hours lost due to equipment problems, 1,035 hours were due to delays resulting from a lack of spare parts. Heat plant outages and minor system modifications resulted in an additional 507 hours of downtime. Table 2.2.1-10 summarizes the downtime during the test period. Once these problems were cleared up, the system has operated very well for over two years as discussed above.

TABLE 2.2.1-10. SUMMARY OF DOWNTIME DURING THE BAHCO SYSTEM TESTING [87]

Category	Downtime Hrs.
Fan	2,766
Auxiliary Equipment	1,088
Heat Plant Outages	388
Modifications	119
Other Items	469
	4,830

In addition to good performance levels in the United States, Japanese FGD installations have also demonstrated high SO_2 removal efficiencies and high reliabilities. Recent reports on Japanese installations have documented system reliabilities of greater than 95 percent. There are several reasons for the good performances being demonstrated in Japan which are discussed briefly below. [88,89]

Generous design - The design of scrubber systems in Japan appear to have generally been approached with a more conservative philosophy than that in the U.S. Specifications provided by users to the scrubber system supplier do not appear to be more detailed than those supplied in the U.S. However, the user expects and, in fact, demands that the scrubber system supplied perform with a reliability compatible with that of the steam generating plant. For example, Electric Power Development Company (EPDC) requires the scrubber system supplier to correct at his expense any process/equipment problems that occur within a year after acceptance of the system. This philosophy has generally resulted in scrubber systems which have required less subsequent modifications and maintenance as a result of fewer operating problems than their U.S. counterparts. The fact that one supplier, Mitsubishi Heavy Industries (MHI), has provided about half of the lime/limestone scrubbing systems in Japan has undoubtedly enhanced the reliability obtained due to MHI's experience.

High limestone utilization - High limestone utilization in the Japanese systems helps prevent mist eliminator plugging and eliminates subsequent downtime.

Lower sulfur fuels - The range of sulfur content of the coal burned in Japanese utility and industrial boilers is lower than that used in U.S. power generating systems. This sulfur content ranges from 0.7 percent at EPDC's Isogo plant to 2.4 percent at Mitsui Aluminum's Omuta Plant with the EPDC Takehara and Taksago plants each averaging around 2.0 percent. However, because of the high ash content and intermediate heating values of the Japanese coals, the SO_2 concentration in the flue gas produced is equivalent to that produced from U.S. coals of somewhat higher sulfur content. For example, the 2.4 percent sulfur coal used in the Omuta plant results in an inlet SO_2 concentration approximately equivalent to a 3.0 percent sulfur Midwest or Eastern coal. Japanese FGD systems have been generally employed with flue gases having SO_2 inlets of 400-2300 ppm. It is significant that this range of inlet sulfur values covers many of the scrubber systems presently applied to U.S. coal-fired utility boilers. However, there is no experience in Japan with the higher sulfur coals currently used in conjunction with many U.S. utility FGD systems. These higher SO_2 content flue gases are generally more inherently difficult to scrub due to mass transfer limitations.

Separate operating crews - The Japanese suppliers and users of FGD systems recognize and accept the fact that scrubbers basically involve chemical processes requiring carefully controlled operation by personnel specifically trained for this purpose. Many utilities including EPDC contract with subsidiary companies to supply scrubber system operating and maintenance services. These specially trained personnel are not rotated into the power plant for other duties as is generally the case with U.S. utilities.

Government surveillance - Japan employs a stringent continuous monitoring and enforcement program to insure that utility and industrial sources are in continuous compliance with environmental regulations. Each prefectural government operates an environmental research center (subsidized by the central government), most of which are directly linked via telemetry systems to automatic monitoring stations located at major emission sources and key ambient sites. The existence of these monitoring systems likely has been instrumental in assuring that emission sources remain in constant compliance since violations would result in fines and/or forced shutdown of the source.

It appears that although the Japanese systems in general operate more reliabily than U.S. systems, the differences between U.S. and Japanese systems are not great. Consequently, it is reasonable to expect the reliability of U.S. systems to increase to Japanese levels in the coming years.

C. Impacts on boiler--The impacts of a lime/limestone FGD system on boiler operations should be small. The most important impacts are: 1)power consumption for running the FGD system's pumps and fans, and 2) possible boiler load reduction during FGD system outages if no bypass is permitted. The amount of power consumed has been estimated to be about 2-4 percent of the net power input to an industrial boiler (see Section 5). Boiler load reduction due to FGD system outages will be the other major impact. However, increasing FGD reliability as demonstrated by the RAFB system will help to reduce these outages to a minimum.

D. Additional maintenance requirements--In all likelihood, additional maintenance will be required to operate a lime/limestone FGD system over that required to operate an industrial boiler alone. The Bahco system located at Rickenbacker A.F.B. has operated with relatively low maintenance requirements once the initial mechanical problems were resolved. Part of the success of this system, however, is probably due to the preventative maintenance program that R.A.F.B. has established which requires one operator/shift which is similar to the Japanese operational plan.[90]

The areas or components in the system that should be given the most attention vary somewhat depending on the design. For most systems they will include the nozzles, headers, strainers, scrubber internals, pump packing and impellers, mist eliminators, fans downstream of the scrubber, agitators, valves, and slurry lines.

2.2.2 Double Alkali Process

The double alkali process encompasses some of the features of lime/limestone wet scrubbing in that a calcium sulfite/sulfate sludge is produced for disposal. While double alkali processes are considered throwaway, a regeneration step is employed to regenerate the active alkali for the SO_2 sorption. The sodium/calcium double alkali process uses a soluble sodium-based alkali for SO_2 sorption and a calcium-based alkali to regenerate the active sodium solution. Although there are several other types of double alkali processes which have been investigated (*e.g.*, ammonia and potassium as the soluble alkali), the sodium/calcium system is the most advanced. The following discussion is limited to this type of system.

2.2.2.1 System Description--

A. Process description--The double, or dual, alkali process uses a soluble sodium-based alkali ($NaOH$, Na_2SO_3, Na_2CO_3 or $NaHCO_3$) to absorb SO_2 from flue gas in a scrubber. The sulfur oxide-rich effluent liquor is reacted with a calcium-based alkali to precipitate calcium sulfite, calcium sulfate, and mixed crystals for disposal. Additionally this reaction regenerates the sodium-based alkali for recycle to the scrubber. The processes developed in the U.S. use lime as the calcium alkali, but other processes developed in Japan and still in the development stage in the U.S. use limestone.

The process can be divided into three principal areas: absorption, regeneration, and solids separation. A simplified flow diagram is given in Figure 2.2.2-1. The principal chemical reactions for a lime regeneration system are described by the following equations.

Absorption

$$2\ NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O \qquad (2.2.2\text{-}1)$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \qquad (2.2.2\text{-}2)$$

Regeneration

$$Ca(OH)_2 + 2NaHSO_3 \rightarrow Na_2SO_3 + CaSO_3 \cdot \tfrac{1}{2}H_2O + 3/2\ H_2O \qquad (2.2.2\text{-}3)$$

$$Ca(OH)_2 + Na_2SO_3 + H_2O \rightarrow 2NaOH + CaSO_3 \cdot \tfrac{1}{2}H_2O \qquad (2.2.2\text{-}4)$$

$$Ca(OH)_2 + Na_2SO_4 + 2H_2O \rightleftarrows 2NaOH + CaSO_4 \cdot 2H_2O \qquad (2.2.2\text{-}5)$$

An important side reaction that also occurs is oxidation:

$$Na_2SO_3 + \tfrac{1}{2}O_2 \rightarrow Na_2SO_4 \qquad (2.2.2\text{-}6)$$

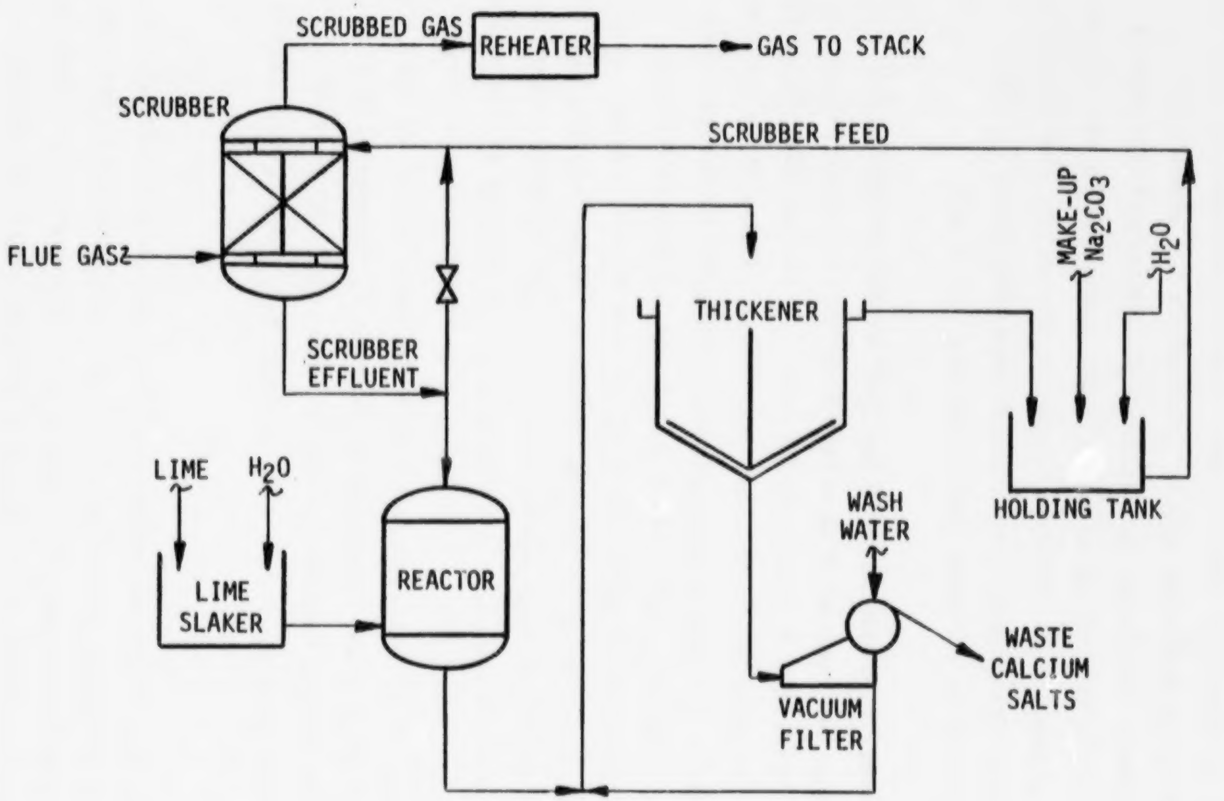


Figure 2.2.2-1. Simplified Flow Diagram for Sodium Double-Alkali Process.[91]

In the absorber, SO_2 is removed from the flue gas by reaction with NaOH and Na_2SO_3, according to Equations 2.2.2-1 and 2.2.2-2. The concentration of active sodium (sodium associated with anions involved in SO_2 absorption reactions, namely sulfite, bisulfite, hydroxide, and carbonate/bicarbonate) may be either concentrated (>.15M) or dilute (<.15M) depending on the system design as described in Section D. Because oxygen is present in the flue gas, oxidation also occurs in the system, according to Equation 2.2.2-6. The scrubbed flue gas may then be reheated before exiting the stack.

Most of the scrubber effluent is recycled back to the scrubber, but a slipstream is withdrawn and reacted with slaked lime in the regeneration reactor according to Reactions 2.2.2-3 and 2.2.2-4. In the process designed by FMC the regeneration reaction occurs in a continuous stirred tank reactor, but in the CEA/ADL design it occurs in a specially designed two-stage reactor system. In the CEA/ADL design, Reactions 2.2.2-3 and 2.2.2-4 both occur; the regenerated solution has a pH of about 11 to 12.5. In the FMC design, however, only Reaction 2.2.2-3 occurs. No free hydroxide is formed, and the pH of the regenerated solution is only about 8.5. The sodium sulfate formed by oxidation of sodium sulfite does not react with lime as readily as the sodium sulfite. The presence of sulfate in the absorber is undesirable in that it converts active sodium to an inactive form, much the same effect as chlorides. Methods of removing sulfates are discussed in Section D, Factors Affecting Performance. As a key design feature currently under study by process developers, this topic is also addressed in Section B.2, Development Status, Recent Improvements.[92]

The regenerated slurry stream, which contains calcium sulfite/sulfate solids, is sent to a thickener where the solids are concentrated. The clear solution overflow from the thickener is stored in a hold tank for return to the absorber, and the underflow containing the calcium sulfite/sulfate solids is further concentrated in a vacuum filter to about 50 percent solids or more. The solids are washed, generally with 1 to 2 displacement washes, to recover sodium salts and then sent to a pond or landfill for disposal. The filtrate and wash water are recycled to the thickener.[93]

B. Development status--

1) Commercial applications - Several vendors in the U.S. have developed double alkali systems. These include Arthur D. Little and Combustion Equipment Associated (CEA/ADL), FMC Corporation, General Motors, Zurn Industries, Neptune Airpol, and Buell/Envirotech. A list of operational and planned double alkali systsms in the U.S. is given in Table 2.2.2-1. In Japan, the double alkali processes developed by Showa Denko and Kureha-Kawasaki (which use limestone as a regenerant) have been applied to about 2450 MW of capacity on oil-fired utility and industrial boilers and sulfuric acid plants. The significant facilities in operation are listed in Table 2.2.2-2. In general, the double alkali process uses technology common to other FGD processes, and the development of double alkali technology benefits from advancements in technology for other soluble sodium scrubbing systems and for lime/limestone systems.

2) Recent improvements - Process improvements and ongoing research have provided potential process alternatives in areas of sulfate removal, forced oxidation and regeneration with limestone. The objectives and current status of each of these areas are outlined below.

Sulfate removal - Various sulfate removal techniques have been investigated to regenerate active alkali from inactive sodium sulfate. While recent process improvements are presented here, a more basic discussion may be found in Section 2.2.1.D.1, Factors Affecting Performance - Design. In dilute systems, with active alkali concentrations less than 0.15 molar, sulfates can be removed by precipitating calcium sulfate with the addition of lime, according to Reaction 2.2.2-5. In a dilute system, however, because of the lower concentration of alkali in the scrubbing liquor, large volumes of scrubbing liquor must be circulated through the absorber to achieve good SO_2 removal.

TABLE 2.2.2-1. FULL SCALE DOUBLE-ALKALI SYSTEMS IN THE U.S.[94,95]

System Operator & Location	System Application	Vendor or Developer	Size (MW, Equivalent)	Active Alkali	Start-up Date
FMC Modesto, CA	Reduction kiln	FMC	10 (gas rate) 30 (regen.)	Conc.	Dec. 1971
General Motors Parma, OH	Coal-fired industrial boilers	General Motors	64	Dilute	March 1974
Caterpillar Tractor Co. Joliet, IL	Coal-fired industrial boilers	Zurn Industries	30	Dilute	Sept. 1974
Firestone Pottstown, PA	Coal-fired industrial boilers (demonstration)	FMC	4	Conc.	Jan. 1975
Gulf Power Co. Scholz Plant, Sneads, FL (Southern Company Services)	Coal-fired utility boiler (prototype)	CEA/ADL	20	Conc.	Feb. 1975[a]
Caterpillar Tractor Co. Mossville, IL	Coal-fired industrial boiler	FMC	70	Conc.	Oct. 1975
Caterpillar Tractor Co. Morton, IL	Coal-fired industrial boiler	Zurn Industries	19	Dilute	Jan. 1978
Caterpillar Tractor Co. East Peoria, IL	Coal-fired industrial boilers	FMC	100	Conc.	April 1978
Caterpillar Tractor Co. Mapleton, IL	Coal-fired industrial boiler	FMC	65	Conc.	(Jan. 1979)[b]
Louisville Gas & Electric Cane Run No. 6	Coal-fired utility boiler	CEA/ADL	277	Conc.	(Dec. 1978)[b]
Central Illinois Public Service, Newton No. 1	Coal-fired utility boiler	Buell/ Envirotech	575	Conc.	(Nov. 1979)[b]
Southern Indiana Gas & Electric, A. B. Brown No. 1	Coal-fired utility boiler	FMC	250	Conc.	(April 1979)
ARCO Polymers Monaca, PA	Coal-fired industrial boilers	FMC	118	Conc.	(June 1980)[b]
Grissom Air Force Base Bunker Hill, IN	Coal-fired industrial boilers	Neptune/ Airpol	15	Conc.	(Sept. 1979)[b]
Chansler Oil Bakersfield, CA	Oil-fired industrial boiler	FMC	35	Conc.	(Mar. 1979)[b]

a. System ceased operation in July 1976
b. Projected Start-up date
c. Size based on a conversion of 2000 SCFM per MW equivalent.

TABLE 2.2.2-2. SUMMARY OF SIGNIFICANT OPERATING FULL SCALE SODIUM/CALCIUM ALKALI SYSTEMS IN JAPAN[96,97]

System Operator & Location	System Application	Vendor or Developer	Size (MW, Equivalent)	Active Alkali	Calcium Sources	Start-up Date
Showa Denko KK Chiba	Oil-fired elec. power boiler	Showa Denko	150	Conc.	Limestone	June 1973
Tohoku Electric Shinsendai Sta.	Oil-fired utility boiler	Kawasaki/ Kureha	150	Conc.	Limestone	Jan. 1974
Nippon Mining, Saganoseki	H_2SO_4 plant	Showa Denko/ Ebara	37	Conc.	Limestone	1973
Toho Zinc, Annaka	H_2SO_4 plant	Showa Denko/ Ebara	43	--	Limestone	1974
Showa Pet. Chem. Kawasaki	Oil fired industrial boiler	Showa Denko	62	--	Limestone	1974
Kanegafuchi Chem. Takasago	Oil-fired industrial boiler	Showa Denko	93	Conc.	Limestone	1974
Poly Plastic Fuji	Oil-fired industrial boiler	Showa Denko/ Ebara	65	Conc.	Limestone	1974
Kyowa Pet. Chem. Yokkaichi	Oil-fired industrial boiler	Showa Denko/ Ebara	46	Conc.	Limestone	1974
Kinuura Utility Nagoya	Oil-fired industrial boiler	Tsukishima	63	Conc.	Lime	1974
Daishuwa Paper Fuji	Oil-fired boiler	Tsukishima	85	Conc.	Lime	1974
Sikoku Electric Power Anan	Oil-fired utility boiler	Kawasaki/ Kureha	450	Conc.	Limestone	1975
Sikoku Electric Power Sakaide	Oil-fired utility boiler	Kawasaki Kureha	450	Conc.	Limestone	1975
Kyushu Electric Power Buzen	2 oil-fired utility boilers	Kawasaki/ Kureha	(2-450) 900	Conc.	Limestone	1977

These large volumes must also be circulated through the regeneration and solid/liquid separation equipment. Capital and operation costs for SO_2 removal would thus be higher for dilute systems.

Sulfate can also be removed by precipitation as a mixed ("double") salt with calcium sulfite as shown in Equation 2.2.2-7.

$$yNa_2SO_4 + xNaHSO_3 + (x+y)Ca(OH)_2 + (z-x)H_2O \rightarrow (x+2y)NaOH + xCaSO_3 \cdot yCaSO_4 \cdot zH_2O \quad (2.2.2\text{-}7)$$

This occurs in concentrated active sodium systems. In this mode, sulfate removal cannot be accomplished by precipitation as gypsum since the high sulfite levels prevent the soluble calcium concentration from reaching a level required to exceed the gypsum solubility product.[93] However, calcium sulfate is precipitated along with calcium sulfite in the regeneration step even though the mother liquor is unsaturated with respect to gypsum. The mixed crystal which results is in effect a solid solution of the two salts. This phenomenon also takes place in lime/limestone wet scrubbing systems.

Another sulfate removal method uses sulfuric acid to dissolve calcium sulfite, increasing the concentration of calcium ions in solution enough to exceed the solubility product of calcium sulfate. The method requires more H_2SO_4 than the stoichiometric amount indicated by Reaction 2.2.2-8, however, and also increases calcium consumption. Thus, it is economically unfavorable in cases of high oxidation or where the gypsum is not marketable. This is the sulfate removal method used in most Japanese double alkali processes.[98]

$$Na_2SO_4 + 2CaSO_3 \cdot \tfrac{1}{2}H_2O + H_2SO_4 + 3H_2O \rightarrow 2NaHSO_3 + 2CaSO_4 \cdot 2H_2O \quad (2.2.2\text{-}8)$$

Forced oxidation - The quality of the solid wastes produced can be improved by forced oxidation of the calcium sulfite to calcium sulfate (gypsum).

Sludges containing a high ratio of calcium sulfate to calcium sulfite are less thixotropic than sludges having a higher proportion of calcium sulfite, and can also be settled, filtered, and dewatered more easily. In addition, calcium sulfate or gypsum has a higher compressive or load-bearing strength and is suitable for landfill disposal. In the Buell-Envirotech process, spent scrubbing liquor is oxidized to further increase the sulfate concentration before it is contacted with lime, thus producing a sludge with higher gypsum content.[99] In Japan, forced oxidation has been used to produce gypsum solid wastes suitable for wallboard or cement production, although this is not a likely market in the U.S. A liquor purge stream may be necessary to prevent the build-up of soluble species (*i.e.*, chloride and magnesium) in systems using forced oxidation since little water is discharged with the solid wastes. This purge stream, however, would be no larger than the amount of water discarded with conventional sludges.

Regeneration with limestone - The double alkali processes developed in the U.S. generally use lime for regeneration, but the use of limestone has been investigated in laboratory tests by A. D. Little.[100] The full-scale processes developed in Japan utilize limestone for regeneration. In these processes, flue gas is scrubbed with a solution of Na_2SO_3. A bleed stream from the scrubber is neutralized with limestone to regenerate the Na_2SO_3 and precipitate calcium sulfite. The calcium sulfite is reslurried, acidified with sulfuric acid to reduce the pH and dissolve calcium sulfite, and oxidized with air to produce gypsum. The gypsum is dewatered to 5-10 percent moisture in a centrifuge. A small slipstream of scrubber effluent is reacted with calcium sulfite and sulfuric acid to remove sulfate. This combination of processing steps makes the process relatively complex and expensive. High levels of oxidation in the scrubber can be tolerated but are uneconomical because of the expense of sulfuric acid addition and the additional limestone required for neutralization.

Magnesium enters as an impurity with the limestone and tends to delay the reaction of limestone with sodium bisulfite in the regeneration step. Limestone is a less expensive raw material than lime but is less reactive so that larger reactors and a more complex reactor system are required. In addition, limestone occurs in many forms which vary in reactivity. Process control may thus be more difficult with limestone than with lime. More solid wastes would also be produced with limestone. Nevertheless, the use of limestone may be an attractive alternative if it can be accommodated with a reasonable degree of increase in process complexity, control requirements, and capital costs.[101]

C. Applicability--Double alkali systems may be somewhat easier to retrofit than lime/limestone systems since, because of the lower liquid-to-gas ratio, smaller equipment is used to contact the flue gas. However, a large amount of land is still required to dispose of the calcium sulfite waste sludge. Future use of double alkali systems on industrial boilers is foreseen since these systems have already been applied successfully to small industrial boilers.[102]

One disadvantage of the double alkali process is the production of large quantities of solid waste. The waste consists primarily of calcium sulfite and generally contains from 30 to 50 weight percent water. Because of high concentration of soluble species in the scrubbing solution, the wastes will also contain soluble solids (such as Na_2SO_3, Na_2SO_4, and NaCl as well as the relatively insoluble calcium salts. The wastes are washed to recover the sodium, and thus the soluble solids content can be reduced to less than 1 weight percent.[103] Even with washing, however, the soluble solids content of double alkali waste solids and the total dissolved solids content of the adherent liquor will probably be somewhat higher than those typically present in lime/limestone wastes.

D. Factors Affecting Performance--

1) Design and Operating Considerations - Potential operating problem areas in double alkali systems that should be considered in system design an operation include oxidation, scale formation, chloride buildup, and sludge quality.

Oxidation - A potential problem in double alkali systems results from the oxidation of sulfite to sulfate. The sulfate thus formed does not take part in SO_2 removal reactions, and is termed "inactive". If the sulfate is not removed from the system, deterioration of SO_2 removal capability or precipitation of sodium sulfate can result.

Theoretically, sulfates can be removed by precipitation as $CaSO_4 \cdot 2H_2O$, according to Equation 2.2.2-5. This reaction, however, does not occur to a significant extent in a solution with a high concentration of sulfite ions. The dissolved calcium ions can bond with either the sulfite ions and precipitate as calcium sulfite or with the sulfate ions and precipitate as calcium sulfate. Since calcium sulfite is less soluble than calcium sulfate, it will precipitate first, leaving the sulfate ions in solution.

If the concentration of sulfite and bisulfite is kept low (active sodium concentration less than 0.15 molar), enough sulfite will be oxidized to sulfate so that there will be more sulfate ions than sulfite ions in solution. If a sufficient excess of sulfate ions is present, the calcium sulfate will be precipitated. A double alkali system in which calcium sulfate precipitates in this manner is classified as a "dilute" system. A system with a higher concentration of sulfite and bisulfite is a "concentrated" system.[10] Because of the lower active alkali concentration in the dilute system, larger volumes of scrubber liquor must be circulated than in a concentrated system to achieve the same level of SO_2 removal. The dissolved calcium concentration will also be higher than that in a concentrated system since calcium sulfate is more soluble than calcium sulfite. Therefore, the dissolved calcium level must build up to a higher level before calcium sulfate is precipitated. The higher calcium level can lead to gypsum scaling in the scrubber, as discussed in the next section. Operation in the dilute mode does allow for easy removal of sulfates, however, and is thus used for applications where the oxidation rate is expected to be high (for example, in systems with low SO_2 and high O_2 concentrations).

Under certain conditions, sulfate is coprecipitated with calcium sulfite, forming a mixed crystal. The amount of sulfate removed in this manner depends on the concentration of sulfate in the reactor liquor. The ratio of

$CaSO_4$ to $CaSO_3$ in the reactor solids appears to be roughly proportional to the ratio of the concentration of $SO_4^=$ to $SO_3^=$ in the reactor liquor.

Scaling - Gypsum and calcium carbonate scaling have occurred in double alkali systems but can be prevented by proper design and operation. Gypsum scaling occurs when sulfate ions formed by oxidation react with calcium ions in the scrubbing liquor, forming gypsum scale. Gypsum scaling is generally not a problem in concentrated systems, since the high concentration of sulfite keeps the calcium ion concentration low. In dilute systems Na_2CO_3 or CO_2 may be added to the liquor to form $CaCO_3$ which will precipitate to reduce the calcium ion concentration before the liquor is returned to the scrubber.

Calcium carbonate scaling can occur in high pH scrubbing liquor when CO_2 absorbed from the flue gas reacts with dissolved calcium ions. This carbonate scaling can be eliminated by pH control in the scrubber. At a scrubber liquor pH below 9 the carbonate/bicarbonate equilibrium tends to limit the concentration of free carbonate ion and thus prevent precipitation of calcium carbonate. If the pH of the regenerated solution is above 9, it should be mixed with the lower pH recycled scrubber liquor before being fed to the absorber to avoid localized high pH areas that could result in carbonate scaling.[105]

Chloride build-up - Another possible problem is the buildup of chlorides in the system. Chlorides are absorbed from the flue gas, and the only mechanism for them to leave the system is in the liquor contained with the solid wastes. The wastes are washed to recover sodium, however, and this washing also recovers chlorides that are then recycled to the absorber. In addition to decreasing the concentration of active alkali in the absorber, high levels of chlorides also can result in stress corrosion.

One possible solution that has been proposed is to use a prescrubber to remove chlorides before the double alkali system. This method has been proposed by Buell-Envirotech in their High Chloride process. The use of a prescrubber with a separate liquor loop, however, could cause water balance problems in the system. Since the evaporation loss would occur in the prescrubber, the only water loss from the double alkali system itself would be the water occluded with the solid waste. This small water loss would not allow addition of enough water for the normal cake washing (more than one displacement wash), demister washing, pump seals, and lime slaking.[106]

Solids quality - Poor settling solids can result from several causes and create problems in thickener and filter or centrifuge operation. ADL has reported that solids settling deteriorates with high calcium sulfate concentrations in the solids. Normally calcium sulfate solids can be dewatered more easily than calcium sulfite. In a concentrated double alkali system, however, the sulfate is incorporated with the calcium sulfite in a mixed crystal rather than as distinct $CaSO_4 \cdot 2H_2O$. High levels of sulfate in the mixed crystal appear to inhibit solids settling. High concentrations of magnesium also inhibit solids settling. In addition, improper control and excess lime feed can make solids settling more difficult.[107]

2) Fuel variations - The importance of fuel characteristics to dual alkali systems is chiefly determined by the following factors:

- Sulfur content
- Chloride content
- Ash alkalinity
- HHV

The effects of these parameters are similar to effects on other FGD systems as discussed in Section 2.2.1.1.D.2. One fuel element which is especially important in dual alkali systems is the potential for sulfite oxidation. Combustion of low-sulfur coals produces conditions which promote oxidation. since the ratio of oxygen to sulfur dioxide in the flue gas is higher than in high sulfur coal applications. Therefore, the sulfur content is a major factor in the selection of design mode and operability of dual alkali systems; *i.e.*, concentrated systems are more suited to application where low oxidation (high sulfur fuel) is expected.[108] Although not a fuel variable, high excess oxygen levels which are commonly found in industrial boilers will also cause high oxidation rates and may affect the design of the FGD system.

Another fuel variable of special importance in dual alkali scrubbing is the chloride content. This species is difficult to purge from the system, but if allowed to build up will significantly reduce the SO_2 absorption capacity of the alkali and/or cause materials problems as discussed previously.

3) Ambient variations - FGD systems are essentially independent of ambient variations. However, as with all wet systems, extreme cold can adversely affect the operation but this effect can be designed for and eliminated by using heat traced lines or by enclosing the FGD system in a building to prevent freezing.

2.2.2.2 System Performance--

A. Emission reductions--Double alkali systems have demonstrated SO_2 removal efficiencies up to 99 percent and have been used to treat gases with inlet SO_2 concentrations as high as 3800 and as low as 250 ppm. Table 2.2.2-3 summarized some of the design features and performance characteristics of operational dual alkali systems.

PEDCo recently summarized some guidelines for achieving high SO_2 removal efficiencies from dual alkali and sodium carbonate systems based on available data.[110] Results of their study recommended the following design and operating practices:

- Upstream flue gas treatment - Prescrubbing with a separate water recirculating system for particulate removal and chloride control for high chloride coal (>0.04 weight percent Cl in the coal).
- Scrubber type - Use of a two-stage tray or packed tower absorber
  --L/G 1.3-2.7 ℓ/m^3 (10-20 gal/1000 acf)
  --absorber pressure drop 15-30 cm (6-12 in.) of water
- pH - recycle liquor pH in 6.0 - 7.0 range
- Reactor tank residence time -
  --In lime regeneration system, approximately 10 minutes with lime utilization ∿90 percent
  --In limestone regeneration, approximately 30 minutes with reactant utilization of 75-85 percent.
- Soda ash (sodium carbonate) addition - to effect carbonate softening and sodium ion makeup

TABLE 2.2.2-3. PERFORMANCE OF DUAL ALKALI FGD FACILITIES IN THE U.S. [109]

Company/Utility:	General Motors	Caterpillar Tractor Co.	Firestone Tire and Rubber Co.	Caterpillar Tractor Co.	Utah Power and Light Co.	Gulf Power Co.
Plant/Unit:	Chevrolet	Joliet Plant	Pottstown Plant	Mossville Plant	Gadsby Station/Unit 3	Scholz, Unit 1
Location:	Parma, Ohio	Joliet, Illinois	Pottstown, Penn.	Mossville, Illinois	Gadsby, Utah	Chattahoochee, Florida
Size of System:	124 m^3/s (262,000 acfm) 32MW	48.8 m^3/s (103,500 acfm) 18MW	6.6 m^3/s (14,000 acfm) 3MW	113 m^3/s (240,000 acfm) 57MW	1.2 m^3/s (2500 acfm) ∿0.6MW	35 m^3/s (75,000 acfm) 20MW
Fuel Characteristics:	Coal 1.5-3.0% sulfur	Coal 4% sulfur	Coal 3% sulfur	Coal 3.2% sulfur	Coal 0.4% sulfur	Coal 3-5% sulfur
Active Alkali Mode:	Dilute	Dilute	Concentrated	Concentrated	Dilute and Concentrated	Concentrated
Startup:	March 1974	September 1974	January 1975	October 1975	1971	February 1975
Status:	Operational	Operational	Operational	Operational	Terminated 1973	Terminated July 1976
SO_2 inlet:	800-1300 ppm	2300 ppm	1000 ppm	--	250-1500 ppm	1800-3800 ppm (dry)
SO_2 outlet:	20-130 ppm	115-350 ppm	100 ppm	--	15-40 ppm	--
SO_2 removal efficiency:	90-99 percent	85-95 percent	90 percent	90 percent	90 percent	95 percent

Some actual test data which document the above guidelines are presented in this section. These specific results are from different test facilities operating under given sets of test conditions, and therefore, may not be generally applicable to all dual alkali FGD systems.

Table 2.2.2-4 summarizes the key operating parameters and results of three test periods for the EPA's sponsored tests at General Motors Parma facility and Figure 2.2.2-2 shows the flow diagram and normal operating conditions. The original literature should be consulted for more details.[111] In general, it was found that despite low SO_2 concentrations in the flue gas (800-1300 ppm by vol.) caused by high excess air, SO_2 removals in the 90 percent range were achieved. This exceeded the average statutory requirement of 60-70 percent SO_2 removal for the coals burned (1.5-3.3 percent S; 11,000-13,500 Btu/lb). Poor SO_2 removal (20-85 percent) in one operating mode was attributed to lower pH in the scrubber which resulted from a modification to the process feed system. Also, calcium sulfite plugging occurred during this operating mode. This mode was, therefore, abandoned for further tests.[114]

The effects of scrubber feed pH and L/G were investigated during pilot testing of Envirotech's double alkali system at the 85 M^3/min (3000 acfm) unit at Gadsby Station.[115,116] The scrubber used for this work was a polysphere unit having perforated plastic balls supported on trays for the actual scrubber packing. Figures 2.2.2-3 and 2.2.2-4 present the results. In general, it was found that SO_2 removals decreased dramatically below pH 6, and that SO_2 removal could be maximized by using multiple trays and by raising the pressure drop across the scrubber. Increasing the liquid flow within a certain range for a given gas flow also increased SO_2 removal.[119]

Performance results have also been reported for the CEA/ADL tests on pilot and prototype levels.[120,121,122] The Scholz prototype unit demonstrated high SO_2 removal capabilities of the dual alkali system when operating in a concentrated mode (0.2-0.4 M active sodium). In over 15 months of operation the average SO_2 removal was 95.5 percent (inlet SO_2 800-2800 ppmv) using both the venturi and absorber. With the venturi alone slightly lower removals

TABLE 2.2.2-4. KEY OPERATING PARAMETERS AND RESULTS FOR INTENSIVE TESTING AT PARMA[112]

	Operating period		
	8/19/74-9/13/74	2/17/75-3/14/75	4/19/76-5/14/76
SO_2 removal	90-95%(rpt. by GM)	65%	90%
Recycle pH	5.5-6	7-7.5	6.0
Scrubber feed location	Top tray	Recycle tank	Recycle line
Lime stoichiometry mole Ca/mole S in cake	1.90	1.65	1.32
Soda ash stoichiometry mole Na_2/mole SO_x removed	-*	0.08	0.12
Solubles loss mole Na_2/mole S in cake	0.11	0.12	0.03
Oxidation mole $CaSO_4$/mole $CaSO_x$ in cake	60%	47%	83%
Ca^{++} in scrubber, ppm	305	465	490
Scaling?	Yes($CaCO_3$)	Yes($CaSO_3$)	No
Na^+ = Total, $\underline{M}$ in scrubber feed	0.58	0.66	0.96†
Na^+ = Active, $\underline{M}$ in scrubber feed	0.087	0.13	0.12
Cake washing?	No	No	Yes
Total solids, wt.%	47.1	41.4	56.0
Soluble solids, wt.% of wet solids	2.4	2.4	1.0
Insoluble solids, wt.% of wet solids	44.7	39.0	55.0
Soluble solids/total solids	0.05	0.06	0.02

*No data taken on SO_2 removal

†Includes ∿0.18$\underline{M}$ Na^+ as NaCl

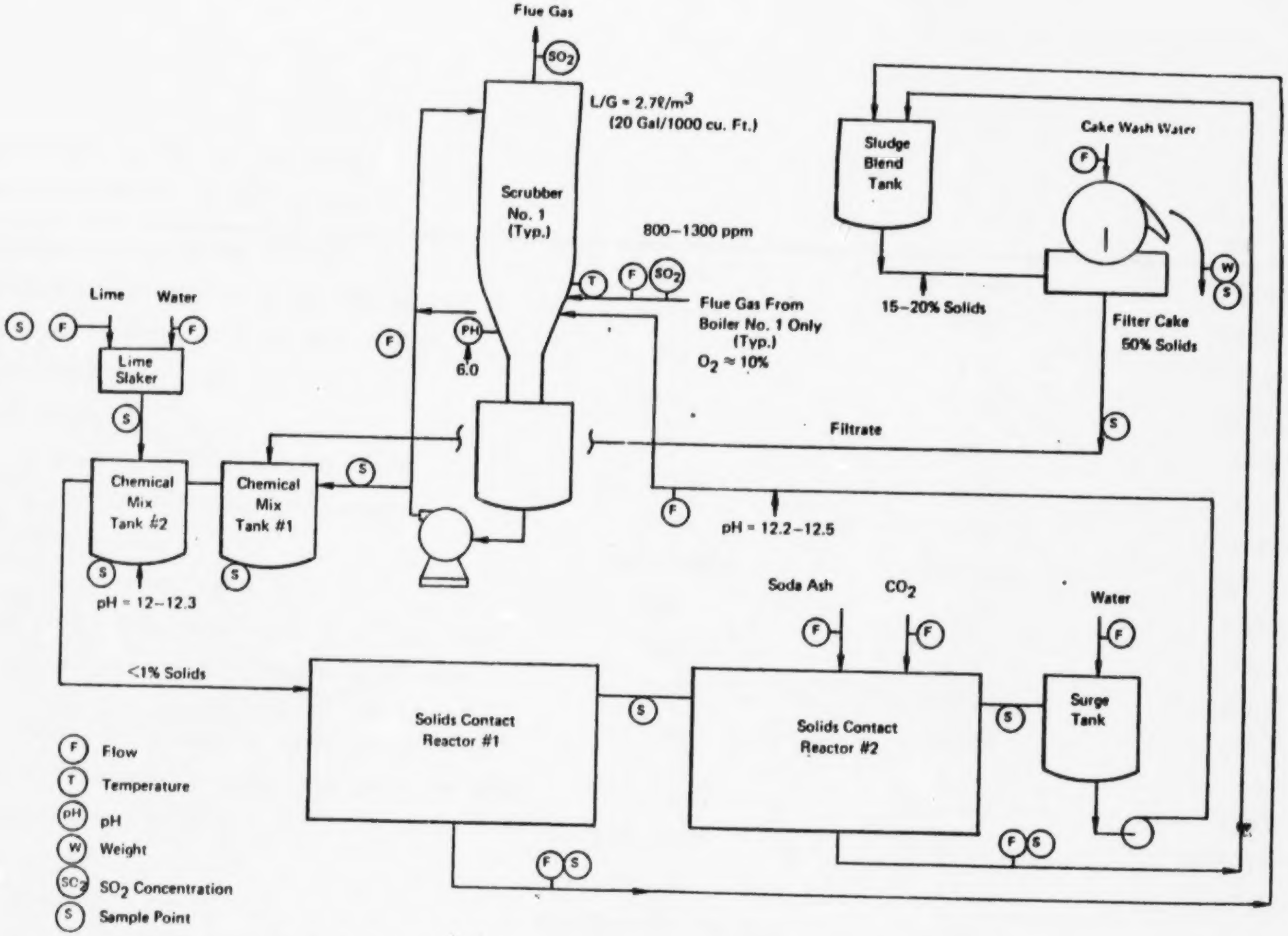


Figure 2.2.2-2. Data collection points and normal operating conditions.[113]

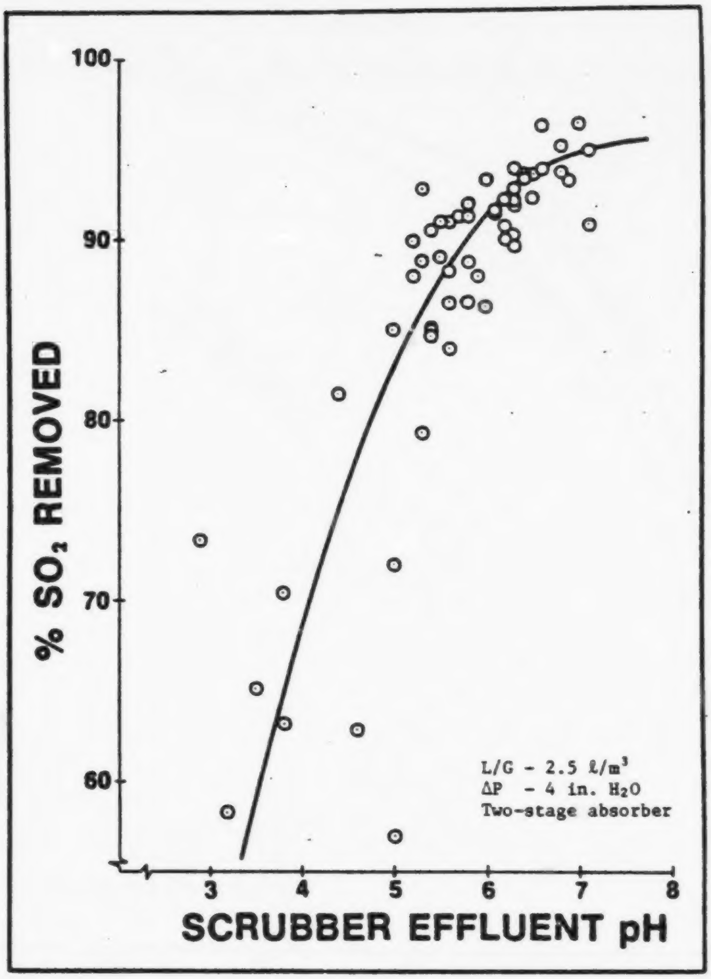


Figure 2.2.2-3. SO_2 removal *versus* scrubber effluent pH for the Envirotech/Gadsby pilot plant with a two-stage absorber.[117]

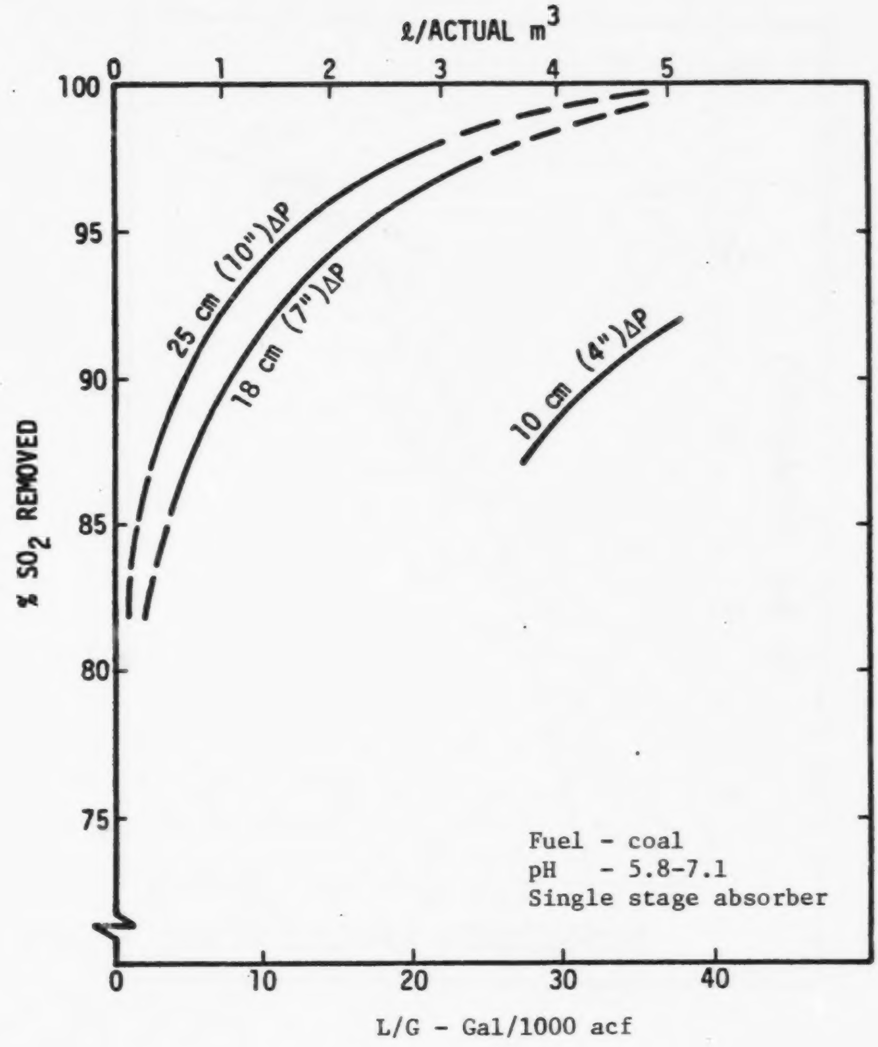


Figure 2.2.2-4. SO_2 removal *versus* L/G ratio for the Envirotech/Gadsby pilot plant with a single stage polysphere absorber.[118]

were obtained, averaging 90.7 percent with SO inlet of 700-1300 ppmv. The chief contorl parameter was pH, with little effect on SO removal found due to highly fluctuating gas flows.

The effects of pH are shown in Figure 2.2.2-5 and 2.2.2-6 for two coal types. The two curves illustrate the range of data for each of the operating configurations evaluated. The operating pH of the scrubber system can be adjusted by changing the sorbent feed rate and/or pH of the regenerated liquor. In concentrated systems the liquid flow requirements for SO_2 removal are low, as illustrated in Figure 2.2.2-7 which shows the required stoichiometry to achieve various SO_2 removal levels over a range of inlet concentrations. To achieve 95 percent removals, L/Gs on the order of 3.3 ℓ/m^3 (25 gal/1000acf) in the venturi and 0.7-0.9 ℓ/m^3 (5-7 gal/10[illegible]acf) in the tray tower were required.

The laboratory and pilot plant work were performed on both dilute and concentrated modes. Results regarding SO_2 removal and oxidation which have recently been published[125] were based on higher scrubber temperatures (60-65°C (140-150°F)) than those normally encountered in conventional boiler flue gas applications (50-54°C (120-130°F)). This would have the effect of decreasing SO_2 removal efficiency (due to elevated SO_2 partial pressures for a given solution) and increasing oxidiation rates. The results, however, are still very useful in generating design data for larger systems. The effects of scrubber feed stoichiometry on SO_2 removal for various SO_2 inlet concentrations are presented in Figure 2.2.2-7. In any range of SO_2 concentration, increasing stoichiometry increased the SO_2 removal, while a higher stoichiometry was required to achieve a given SO_2 removal in the lower inlet SO_2 range than in the higher inlet SO_2 range.

B. Availability/reliability--Since 1973 double alkali systems have been started up on three large utility boilers and several industrial boilers in Japan. Although detailed operating details are not available, few

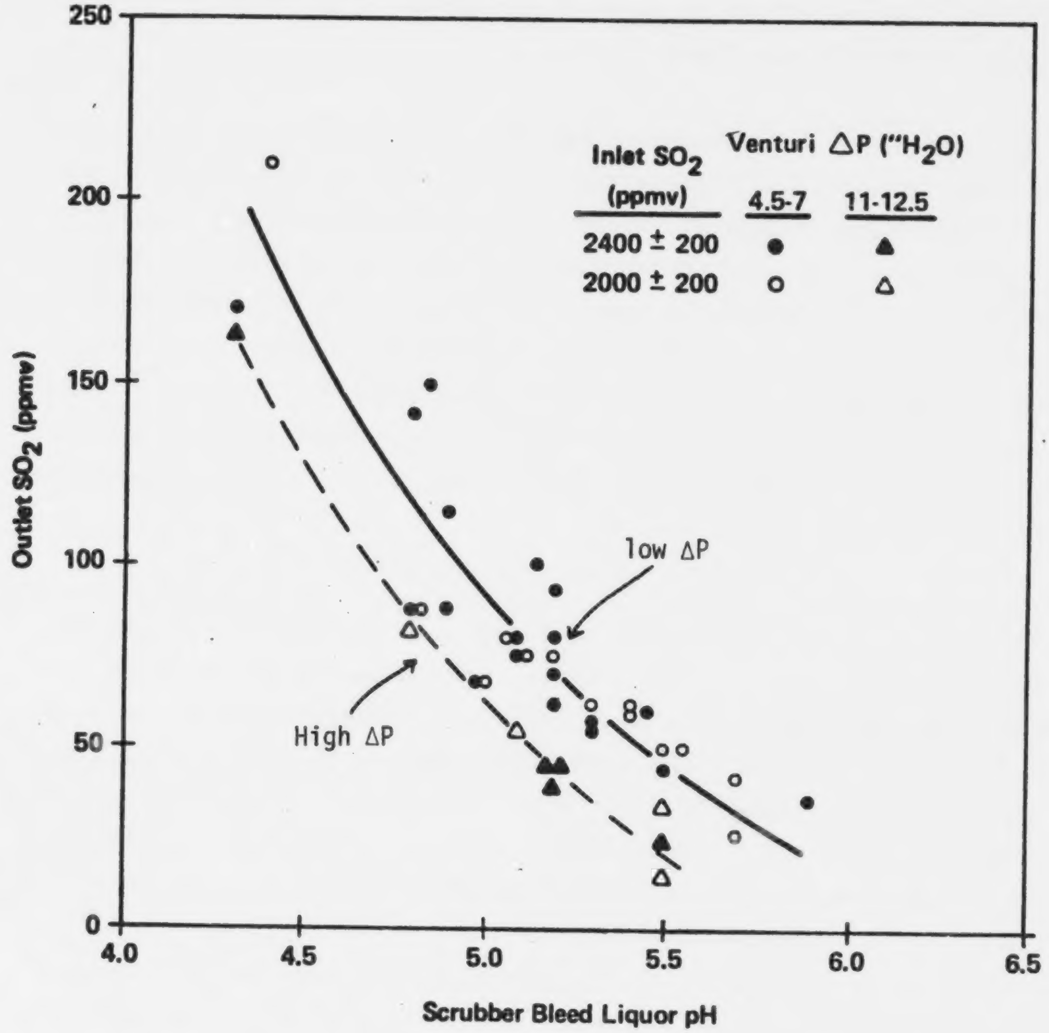


Figure 2.2.2-5. CEA/ADL Dual Alkali Process - SO_2 removal as a function of pH - high-sulfur coal.[123]

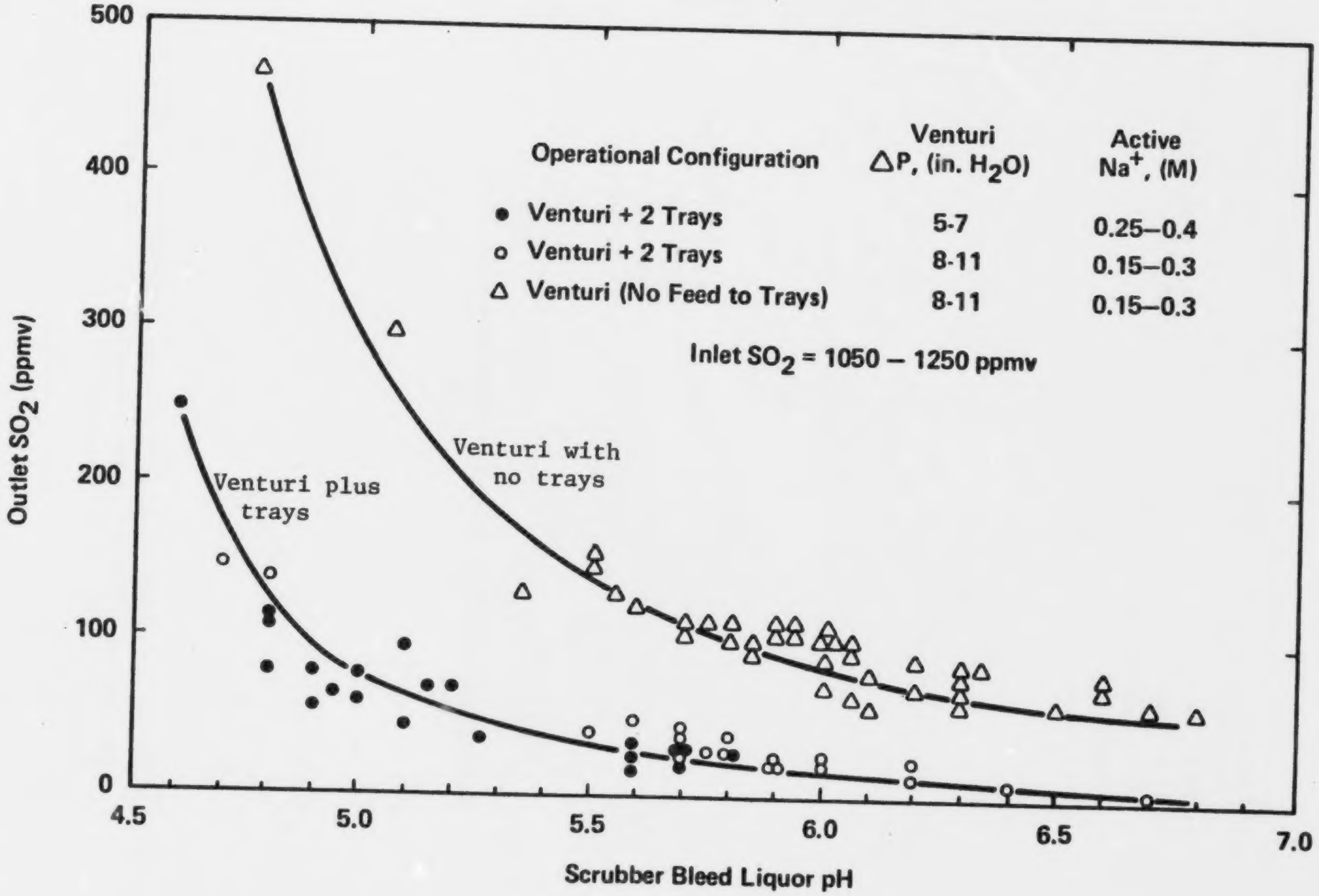


Figure 2.2.2-6. CEA/ADL Dual Alkali Process - SO_2 removal as a function of pH - low/medium-sulfur coal.[124]

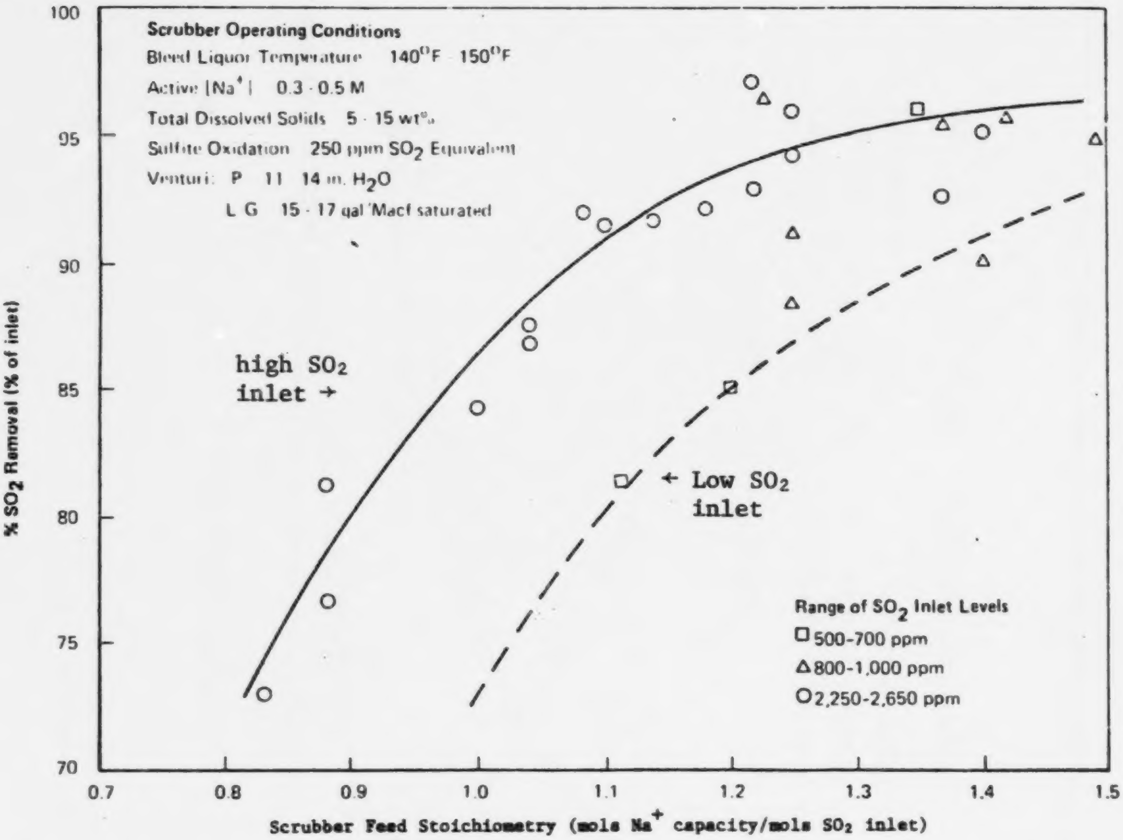


Figure 2.2.2-7. Effect of feed stoichiometry on removal efficiency in the Venturi/2-tray tower absorber for the EPA-ADL double-alkali pilot program.[126]

problems have been encountered.[127] Scaling and corrosion, while not rare in any FGD operation, are reportedly controlled so that they do not hinder overall plant operation.[128] Operabilities for the Shikoku Electric Sakaide plant and the Showa Denko Ichihara plant were reported to be 95 and 98 percent, respectively, with corresponding SO_2 removal efficiencies of 95 and 90-95 percent.[129] Operability is defined as the ratio of the FGD system's operating hours to the scheduled operating hours of the gas source, normally 11 months continuous operation with a one month shutdown for utility boilers.

Since there are few double alkali systems with long operating histories in the U.S., it is difficult to predict the long-term reliability of a system for a given industrial boiler application. The largest operational double alkali system in the U.S. is operated by Caterpillar Tractor in Mossville, Illinois. This system, which was designed by FMC, removes the full load of fly ash from the flue gas. Several problems have been experienced including plugging of the demister (which has occurred several times), difficulty in controlling the liquor flow rate to the absorber, difficulty in dewatering fly ash, and erosion problems.[130] Most of these problems, however, are apparently due to the large amounts of fly ash in the system. Thus, a system installed with upstream particulate removal should be more reliable than the Caterpillar installation.

The 3 MW demonstration unit designed by FMC for Firestone Tire and Rubber, Pottstown, Pennsylvania, had an overall reliability (based on boiler operation) of about 95 percent when firing with oil (93.5 percent the first year) and 88 percent when firing with coal. Figure 2.2.2-8 illustrates the reliability of this system over the last four years. Nothing was spared in this small installation, including pumps, which accounted for 15 percent of the total system downtime the first year. Most of the remaining downtime the first year was due to thickener pluggage and problems with the cake conveyor, fan, lime feeder, and spray nozzles, with less downtime due to instrumentation and control valves. During the period of coal

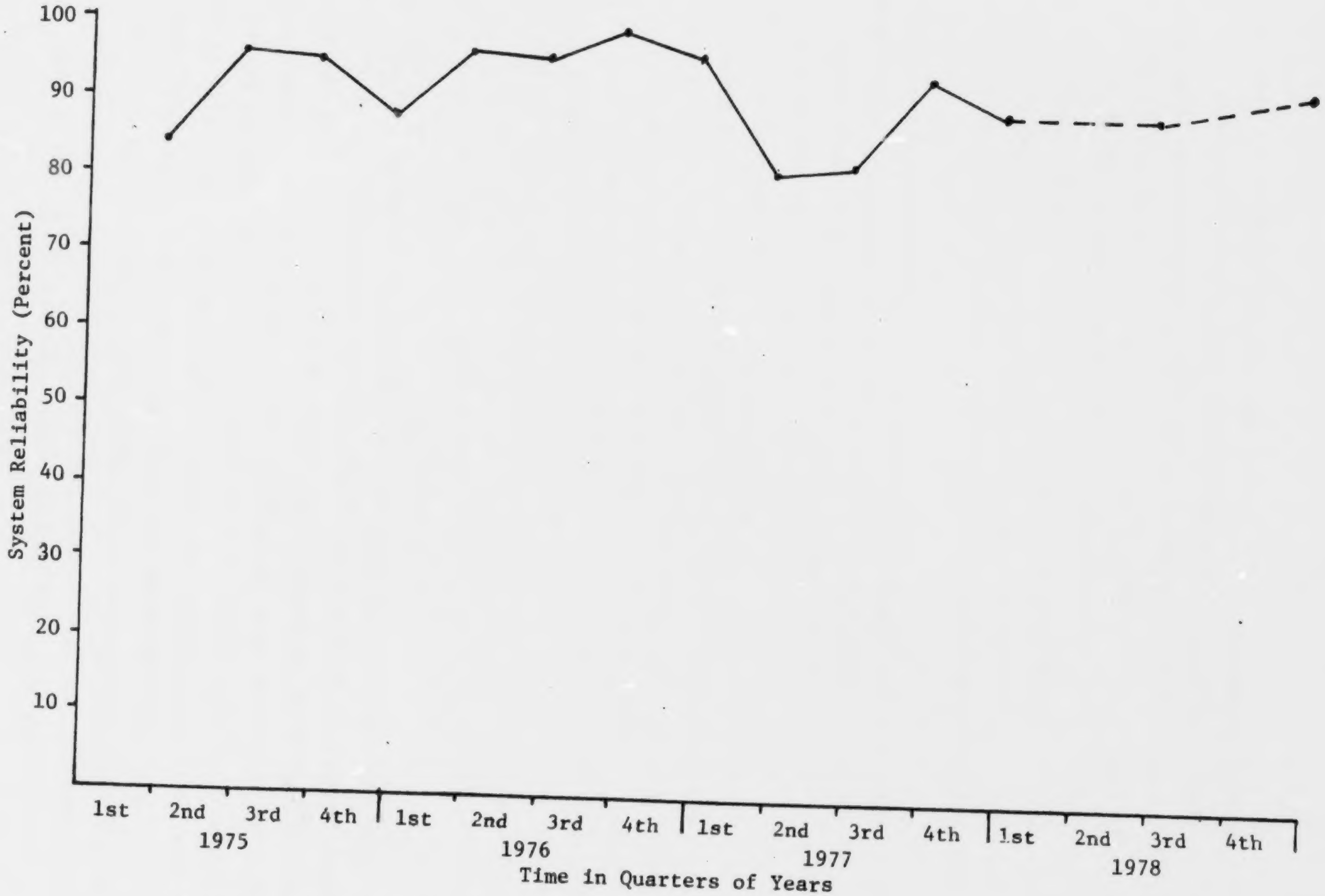


Figure 2.2.2-8. Performance trends for the Firestone double alkali system[131]

firing, most of the downtime was due to fly ash, which the system was not designed to handle. Erosion caused by fly ash has ruined the fan and damaged the venturi nozzles. Firestone has been satisfied with the reliability demonstrated by the system. They are investingating the possibility of converting their boilers to coal, using the double alkali process to control SO_2 emissions.[132]

The oldest full-scale double alkali system in the U.S., at FMC in Modesto, California, treats gas from reduction kilns. The gas has a high SO_2 content, up to 8000 ppm SO_2. The system reportedly has had an availability over 95 percent since it started up in December 1971. In 1977 the scrubbing system reportedly did not cause any kiln downtime (100 percent operability). The scrubber (a packed tower) is flushed with water for cleaning during normal plant maintenance shutdowns, which occur for several days about 3 times a year. The gas to be cleaned contains calcium, which has led to scaling in the scrubber packing. The packing has been changed twice and cleaned with acid four times since 1971, but no scaling has been noted in the year since the recirculation rate was increased.[133]

The purpose of the 20 MW prototype unit at Scholz, designed by CEA/ADL, was to characterize and evaluate the performance of the double alkali process. It was not intended as a demonstration unit nor a test of the ultimate reliability possible from a full-scale application. Nevertheless, the overall operative record of the system over the 17 months of operation from initial startup through completion of the test program was impressive, including delays in receiving replacement parts and in awaiting repair of the boiler air preheater. Most of the downtime occurred between operating periods; during the operating periods the availability averaged about 90 percent. The various parameters actually achieved by this unit are shown in Table 2.2.2-5.

As reported by Rush and Edwards[135] the process operability (ease of operation) was excellent in all respects. The system was successfully

TABLE 2.2.2-5. CEA/ADL DUAL ALKALI PROCESS VIABILITY PARAMETERS[134]
(February 1975-July 1976)

Parameter	Value %	Total time (hours)	
		Numerator	Denominator
Availability*	78	9,679	12,376
Reliability†	80	7,128	8,911
Operability‡	70	7,128	10,172
Utilization Factor§	58	7,128	12,376
Parameter 5#	95	9,679	10,172

*Availability: Hours the FGD system was available for operation (whether operated or not) divided by hours in the period, expressed as a percentage.

†Reliability: Hours the FGD system was operated, divided by the hours the FGD system was called upon to operate, expressed as a percentage.

‡Operability: Hours the FGD system was operated divided by boiler operating hours available to the process in this period, expressed as a percentage.

§Utilization Factor: Hours that the FGD system operated divided by total hours in this period, expressed as a percentage.

#Parameter 5: Hours the FGD system was available for operation (whether operated or not) divided by the hours the boiler was available to the process, expressed as a percentage.

operated over a range of widely fluctuating inlet flue gas sulfur dioxide levels (1.6 to 4 wt% sulfur fuel), oxygen concentrations, and flow rates with little or no change in the sulfur dioxide removal efficiency, waste cake properties, or lime utilization. It demonstrated an ability for continuous operation with large and frequent variations in pH in both the scrubber circuit (4 to 7) and the regeneration/solids dewatering section (6.5 to 13).

The most impressive aspect of system operation was its resistance to short-term upsets. Low soluble calcium levels throughout the system, even during most upset conditions, resulted in a low potential for calcium/sulfur salt precipitation, particularly in the scrubber circuit. In addition, operation for over 4500 hours without mist eliminator washing and without any mist eliminator scaling confirmed the viability of system operation without a mist eliminator wash.

The dual alkali process does have a limitation in that there is a minimum level of sulfur content in the fuel below which it cannot be successfully operated in the concentrated mode due to the potential for gypsum scaling. This lower limit is a function of the rate of oxidation in the system and is therefore dependent not only on the sulfur content in the fuel, but also the level of oxygen in the flue gas. For typical pulverized-coal-fired boilers, the design of a concentrated-mode dual alkali system for use on flue gases produced from the combustion of coal containing between 1 and 2 wt% sulfur will have to be considered carefully on a case-by-case basis. For fuels containing less than 1 wt% sulfur, a concentrated-mode dual alkali system cannot be operated at the excess air levels that are typical for pulverized-coal-fired boilers without an intentional purge of sodium sulfate, either in the filter cake or as a separate purge stream. However, above 2 wt% sulfur in the fuel, and in many cases for coal containing between 1 and 2 wt% sulfur, the operation of the system is excellent. In fact, overall system operation improves significantly as higher sulfur content fuel is utilized.[135]

It is difficult to generalize with respect to the long-term reliability potential of any mechanical/chemical process. However, the results at Scholz indicate that the overall performance of a properly designed and operated dual alkali system can be superior to that of direct lime and limestone systems because:[136]

- As discussed above, the system is highly resistant to upset. The potential for calcium/sulfur salt precipitation (scaling) is eliminated by the high sulfite levels in solution, except in extreme upset conditions.
- The handling of slurries in the absorption section is completely eliminated.
- The most important control parameter--pH--has a wide, acceptable range of operation, and unlike lime and limestone systems, has no effect on scaling up to values of 12.5 to 13 in the reactor system.

The General Motors' double alkali system at Parma, Ohio, was started up in February 1974. The dilute process was the subject of an EPA-sponsored test program from August 1974 to May 1976, during which significant improvements in both process and mechanical performance were observed. Test objectives included determination of SO_2 removal capability, process reliability, sulfate control, waste characteristics, degree of closed loop operation, lime and other chemicals utilization, and material balances. The test program involved three one month intensive tests and 18 months of lower level nonintensive testing. Numerous modes of operation were investigated, necessitating shutdowns for equipment changes and modifications. There were, however, several shutdowns due to equipment or other operational malfunctions as described in detail in the literature.[137] The overall results and conclusions regarding system reliability were that, in part due to its then developmental status, it was judged "not yet proven over an extended period to be a commercially viable process for SO_2 control."[138] However, significant improvements observed during the last four week test period (April-May 1976) were thought to indicate that the system had capability for long-term reliability. The total scrubber availability to the boilers over the whole test program was 77.9 percent, excluding four long-term planned shutdowns for system modifications.[139]

During the two year period since the end of the EPA study the Parma FGD system has continued to operate, experiencing varying operability indexes (1.25-100 percent). Figure 2.2.2-9 presents operability data from the General Motors Parma system over the last four years. Shutdowns have been for annual scheduled overhauls, severe winter conditions, low boiler loads during summer months, and in some cases mechanical problems. The most frequent problem was continued solids buildup in the mix tank. To resolve this problem the mix tank agitators have been totally redesigned and replaced by high-carbon steel, propeller types.[140]

C. Impacts on boiler--The major impacts of a dual alkali FGD system on boiler operations are similar to those of a lime/limestone system; *i.e.*, 1) power consumption for running the FGD system's pumps and fans and 2) possible boiler load reduction during FGD system outages if no bypass is permitted. The power consumption for the CEA/ADL prototype unit at Scholz was reported to be 2.5-3.0 percent of the unit's generating capacity with the system operating at full fan and at full venturi and absorber liquid recirculation capacity. Correcting for the excess fan and pump capacity, the power consumed by the equipment actually required in the application (tray tower at an L/G of 0.7-1.3 ℓ/m^3 or 5-10 gal/1000 ft^3) is approximately 1-1.5 percent of the design generating capacity.[141] Although an industrial boiler generally produces steam instead of electricity, the proportion of boiler product usage by the dual alkali system should be about the same assuming that steam can be used to drive the pumps and fans.

Boiler load reduction due to FGD system outages will be the other major impact. The relatively high system reliabilities exhibited by existing double alkali processes will help to reduce these outages to a minimum.

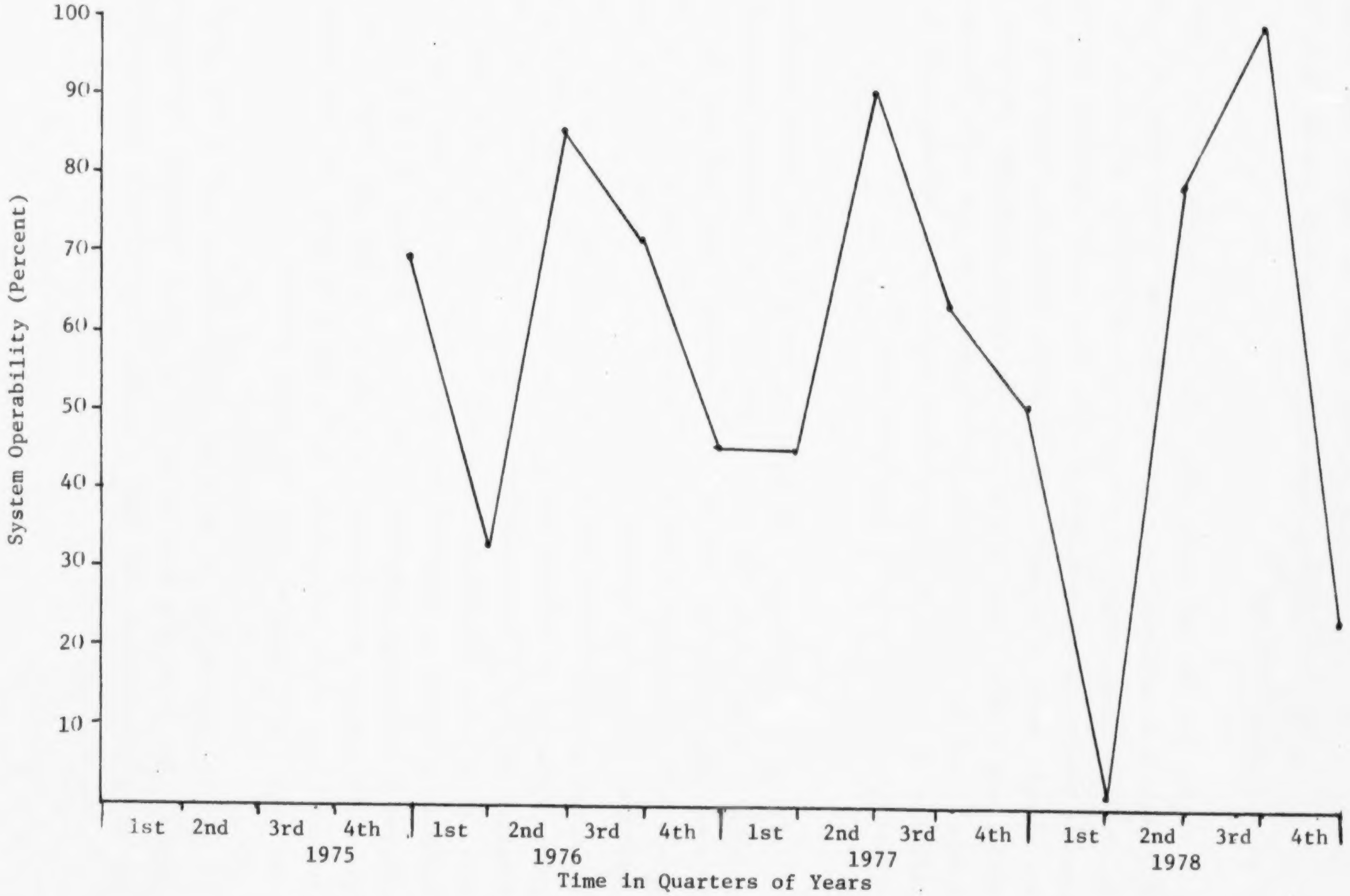


Figure 2.2.2-9. Performance trends for the GM Parma double alkali system[140]

D. Additional maintenance requirements--Operation of a dual alkali system increases the total maintenance requirement of an industrial boiler facility. Manpower requirements for the GM double alkali process is reported to be 1.4 men/shift for direct operation and 0.5 man/shift for maintenance.[142] FMC claims that under most conditions one boiler house operator is sufficient to operate the FGD system. In comparison to direct lime/limestone scrubbing, the maintenance requirement should be less because the scaling potential is minimized and a solution rather than a slurry is used for scrubbing.[143]

2.2.3 Wellman-Lord Sulfite Scrubbing Process

2.2.3.1 System Description--

A. System--The Wellman-Lord Sulfite Scrubbing Process marketed by Davy Powergas is based on the ability of sodium sulfite solution to absorb SO_2, thus forming a solution of sodium bisulfite which can be thermally regenerated. It is a regenerable process and is presently being commercially employed on a large scale. A concentrated stream of SO_2 is produced which can be processed to elemental sulfur, sulfuric acid, or liquid SO_2. A by-product purge of sodium sulfate is produced as the result of sulfite oxidation. Antioxidants have been used to reduce the sulfate purge rate, but have been found to be uneconomical.

A simplified process flow sheet appears in Figure 2.2.3-1. The Wellman-Lord Process consists of five basic processing steps:

1) Gas Pretreatment
2) Absorption
3) Purge Treatment
4) Regeneration
5) SO_2 Conversion

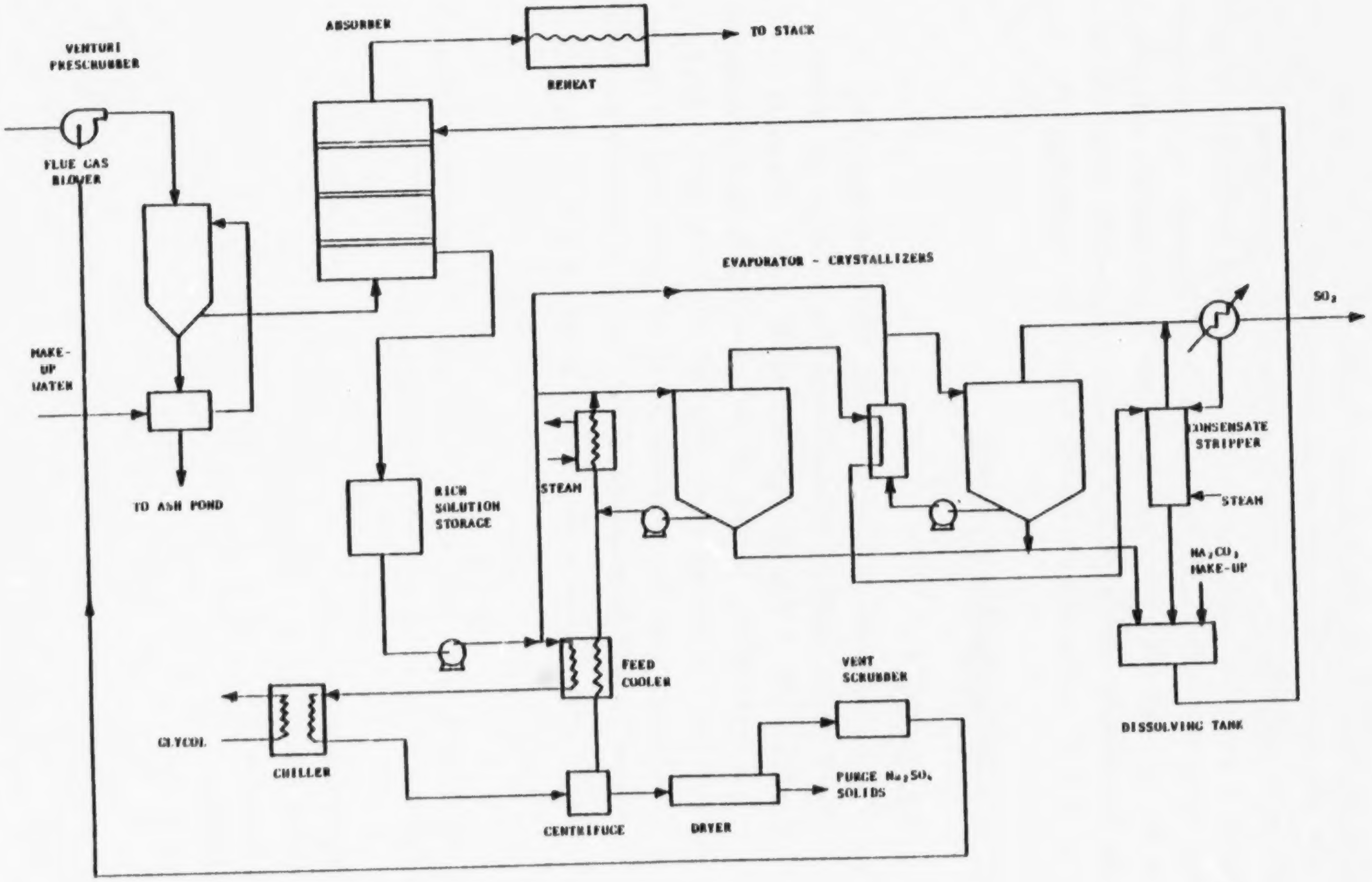


Figure 2.2.3-1. Process flow diagram Wellman-Lord Process.[144]

No unusual or unique equipment is used in any of these areas with the possible exception of the SO_2 conversion step which is licensed technology. The gas pretreatment and absorption sections are essentially the same as those found in most aqueous scrubbing systems.

1) Gas pretreatment - The flue gas to be treated is taken downstream of the electrostatic precipitator. This gas, which is at a temperature of about 300°F, is passed through a venturi or tray type prescrubber where it is cooled to around 130°F and humidified. The venturi scrubber is preferred because it removes 70 to 80 percent of any remaining fly ash and 95 to 99 percent of the chlorides. A tray-type prescrubber is satisfactory for cooling and humidifying low pressure drop, but is less efficient than a venturi for fly ash and chlorides removal. Humidification of flue gas in the prescrubber prevents evaporation of excessive amounts of water in the absorber.

A well-designed prescrubber can remove up to 99 percent of all chlorides in the flue gas.[145] This should help maintain a low level of chloride in the scrubbing liquor, and reduce the potential for stress corrosion. Fly ash and other solids collected by the prescrubber are pumped to the ash disposal pond as about 5 percent slurry. When absorption of $HC\ell$, SO_2, and SO_3 causes the slurry water to become too acidic, the slurry is neutralized with lime before it is pumped to disposal.

2) Absorption - Humidified gas from the prescrubber is passed through the absorption tower where the SO_2 level is reduced by at least 90 percent. The cleaned gas may be reheated by heat exchange with high-pressure steam and exhausted to the atmosphere. There are various alternatives to this method of reheating the gas, but use of steam allows coal to be used indirectly rather than premium fuels such as oil or natural gas.

Absorption of SO_2 proceeds according to Equation 2.2.3-1.

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \qquad (2.2.3\text{-}1)$$

Makeup sodium carbonate also reacts with SO_2 in the absorber to form sodium sulfite by Equation 2.2.3-2.

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2 \qquad (2.2.3\text{-}2)$$

A very important side reaction, which will be discussed in detail later, is the oxidation of sulfite to sulfate by oxygen in the flue gas as in Equation 2.2.3-3.

$$Na_2SO_3 + \tfrac{1}{2}O_2 \rightarrow Na_2SO_4 \qquad (2.2.3\text{-}3)$$

Some sodium sulfate is also formed by absorption of SO_3 from the flue gas:

$$2Na_2SO_3 + SO_3 + H_2O \rightarrow 2NaHSO_3 + Na_2SO_4 \qquad (2.2.3\text{-}4)$$

Davy Powergas offers two types of absorption units, a packed tower for small volume applications and a valve tray tower for large volume applications. The valve tray unit is generally built in a square configuration and includes three to five trays depending on the inlet SO_2 concentration and the degree of desulfurization. Because of the large capacity of concentrated sodium sulfite solution to absorb SO_2, the feed liquor flow is fairly small and recirculation is practiced on each stage to maintain good hydraulic characteristics over the trays. With the recirculation, the L/G ratio is kept at about 3 gal/mscf, per tray. Superficial gas velocity in the absorber is about 10 feet per second.

Scaling and plugging problems in the absorber are virtually eliminated because:

- the prescrubber removes solids,
- the scrubber solution is clear, and
- the absorption product, sodium bisulfite, is more soluble than sodium sulfite.

3) Purge Treatment - About 15 percent of the absorber product liquor is drawn off to the purge treatment area for separation of sodium sulfate. Feed liquid is cooled by heat exchange with treated liquid. In a chiller-crystallizer the purge is further cooled to about 32°F and a mixture of sodium sulfate and sulfite is crystallized out. The slurry is put through a centrifuge to produce a 40 percent solids cake and the cake is dried by steam. A small liquid purge stream is also drawn off the evaporators in the regeneration section and added to the cake to remove some of the thiosulfate formed in the system. The crystalline product is a mixture of anhydrous sodium sulfate (70 percent) and sodium sulfite (30 percent), plus small amounts of thiosulfates, pyrosulfites, and chlorides. The vent gases from the dryer are scrubbed to remove dust and recycled to the main flue gas stream. The centrate is heated by passing through the feed cooler and is returned to the product liquor stream entering the evaporator loop. Refrigeration for the chiller-crystallizer is provided by an ethylene glycol system.

4) Regeneration - The regeneration section consists of a set of double effect, forced circulation evaporators, condensers, a condensate stripper, and a dump dissolving tank. Regeneration of sodium sulfite is performed by simply reversing the absorption reaction through addition of heat as shown in reaction 2.2.3-5.

$$2NaHSO_3 \rightarrow Na_2SO_3 + H_2O + SO_2 \qquad (2.2.3\text{-}5)$$

However, higher temperatures also increase the formation of thiosulfate by the following disproportionation reactions.

$$6NaHSO_3 \rightarrow 2Na_2SO_4 + Na_2S_2O_3 + 2SO_2 + 3H_2O \qquad (2.2.3\text{-}6)$$

$$2NaHSO_3 + 2Na_2SO_3 \rightarrow 2Na_2SO_4 + Na_2S_2O_3 + H_2O \qquad (2.2.3\text{-}7)$$

As depicted in Figure 2.2.3-1, the combined stream of absorber product liquor and purge centrate is split between the two evaporator effects, each of which operates under a vacuum, with 55 percent going to the first effect and 45 percent going to the second effect. The first effect operates at 200°F and is heated with low pressure steam by way of an external shell and tube exchanger. The SO_2 and water vapor driven off in the overhead from the first effect are used to heat the second effect which operates at about 170°F. In each effect the undissolved solids content of the recirculated liquor is about 45 percent, primarily sodium sulfite. Operating at such a high solids concentration eliminates the need to centrifuge the product stream going to the dissolving tank. The regeneration reaction is limited by the equilibrium concentration of sulfite ion in solution. Fortunately, since sodium sulfite is less soluble than sodium bisulfite, it is continuously removed from solution by crystallization, driving the reaction forward.[146]

The SO_2 and water vapor from the evaporators is subjected to partial condensation to remove most of the water and thus concentrate the SO_2. The condensate contains several hundred ppm of dissolved SO_2 and is steam stripped to lower these values. Stripped condensate is returned to the dump dissolving tank along with a small amount of makeup water and sodium carbonate. This mixture is then agitated with the sodium sulfite slurry from the evaporators to provide absorber feed. The SO_2 stream exiting the condenser contains only 5 to 10 percent water. It is compressed and transferred to the SO_2 conversion section.

5) SO_2 Conversion - One of three product options may be selected for conversion of gaseous SO_2 to useful chemicals. Sulfur dioxide may be:

- compressed and liquified for sale,
- catalytically oxidized for production of sulfuric acid (Contact Process), and
- reduced to elemental sulfur.

These options are listed in order of increasing processing cost.

The questions of what to do with the SO_2 product stream will have to be answered separately for each individual site. But for production of sulfur to be a viable alternative, a process able to use a reductant other than methane will often be necessary. Current Wellman-Lord sulfur producing systems employ the Allied Chemical sulfur production unit which uses methane as the reductant. In some isolated cases, methane may be readily available but in general the most reliable sources of reducing agent will be coal or coke. Coal might be used in solid form as in Foster Wheeler's RESOX process or as H_2/CO producer gas. Unfortunately, this whole area is a source of uncertainty because to date, only the Allied Chemical process using methane has been demonstrated on a large scale. The use of reducing gas from coal gasification in the Allied Process is still in the relatively early developmental stage.

B. Status of development--The Wellman-Lord Process was first conceived in 1966 and a pilot plant using potassium sulfite rather than sodium sulfite was operated in 1967. The problems encountered with scaling and high steam consumption on this original pilot induced a switch to the sodium system. The potassium bisulfite product is less soluble than potassium sulfite, while sodium bisulfite is more soluble than sodium sulfite. Thus, in a sodium system, as SO_2 is absorbed and water evaporates, the saturation point will not be exceeded because the product of absorption, sodium bisulfite, is more soluble than the reactant, sodium sulfite.

The first sodium sulfite system was installed for Olin Corporation in 1970 to treat a 45,000 scfm stream of acid plant tail gas. Acid plant tail gas differs from industrial boiler flue gas in that it generally contains a higher SO_2 concentration (up to 10,000 ppm) and a lower oxygen concentration. Despite some initial difficulties, the plant operated successfully. Subsequently, the process was applied to an oil-fired industrial boiler of the Japan Synthetic Rubber Company in 1971. Again, the plant was operated successfully after some initial difficulty. This unit has since been able to achieve better than 90 percent removal of SO_2 with an onstream factor of better than 97 percent.

Wellman-Lord systems currently account for about 4 percent of the total USA operating and planned utility FGD capacity,[147] and about 10 percent of total Japanese capacity.[148] Although the fuel base is not specified in these statistics, it is safe to say that the U.S. systems are predominately coal-fired, and the Japanese, oil-fired.[149]

Tables 2.2.3-1 and 2.2.3-2 summarize the operating Wellman-Lord systems in the U.S. and Japan, and Table 2.2.3-3 summarizes the systems planned for U.S. operation.

Table 2.2.3-4 gives worldwide distribution of Wellman-Lord units. Of these 31 installations, two of the most significant are the unit in operation since 1973 for Chubu Electric Power's Nagoya Station and the unit at Northern Indiana Public Service Company's Dean H. Mitchell Station. The Chubu plant is significant because it is the largest Wellman-Lord system with a fairly long operating history. The application is a 220 MW, oil-fired peaking plant. This Wellman-Lord Unit has been highly successful, achieving 90 percent removal of SO_2 consistently with a high on-stream factor. The

TABLE 2.2.3-1. SUMMARY OF OPERATING WELLMAN-LORD SYSTEMS IN THE U.S.[150]

Company/location	Completion date	Feed gas origin	Gas flow 1000 Nm^3/hr	Gas flow (scfm)	Design SO_2 concentration, ppm	Disposition of SO_2
Olin Chemical* Paulsboro, New Jersey	July 1970	Sulfuric acid plant	76	(45,000)	in 6,000 out 500	Recycle to acid plant
Std. Oil of California El Segundo, California	September 1972	Claus plants	51	(30,240)	in 10,000 out 250	Recycle to Claus plant
Allied Chemical Calumet, Illinois	November 1972	Sulfuric acid plants	51	(29,850)	in 2,700 out 250	Recycle to acid plants
Olin Chemical Curtis Bay, Maryland	May 1973	Sulfuric acid plants	133	(78,046)	in 4,000 out 250	Recycle to acid plants
Std. Oil of California Richmond, California	August 1974	Claus plant	51	(30,000)	in 10,000 out 250	Recycle to Claus plant
St. Oil of California El Segundo, California	January 1975	Claus plant	51	(30,000)	in 10,000 out 250	Recycle to Claus plant
Northern Indiana Public Service Gary, Indiana	December 1976**	Coal-fired boiler (115 MW)	527	(310,000)	in 2,200 out 200	Elemental sulfur
Public Service Co. of New Mexico Waterflow, New Mexico	November 1978	Coal-fired boiler (314 MW)	—	—	90% removal	Elemental sulfur
Public Service Co. of New Mexico Waterflow, New Mexico	November 1978	Coal-fired boiler (306 MW)	—	—	90% removal	Elemental sulfur

* Plant operation suspended as of January 1, 1976.

**First integrated operation of the plant was in December, 1976; however, the Wellman-Lord system began operations in July 1976, before completion of the Allied sulfur recovery unit.

TABLE 2.2.3-2. SUMMARY OF OPERATING WELLMAN-LORD SYSTEMS IN JAPAN[151]

Company/location	Completion date	Feed gas origin	Gas flow, $1000\ m^3/hr$	Gas flow, (scfm)	SO_2 concentration, ppm	Disposition of SO_2
Japan Synthetic Rubber Chiba	August 1971	Oil-fired boilers	200	(124,000)	in 2,100	Sulfuric acid
Toa Nenryo, Kawasaki	August 1971	Claus plants	67	(41,000)	in 6,500 out 200	Recycle to Claus plants
Chubu Electric Power, Nagoya	May 1973	Oil-fired boiler (220 MW)	620	(390,000)	in 1,600 out 150	Sulfuric acid
Sumitomo Chemical, Sodeguara	November 1973	Oil-fired boiler	360	(225,000)	in 1,550 out 250	Sulfuric acid
Japan Synthetic Rubber Yokkaichi	December 1973	Steam boiler (2% S fuel oil)	450	(280,000)	in 1,000 out 100	Sulfuric acid
Kashima Oil, Kashima	February 1974	Claus plants	30	(20,200)	in 11,000 out 200	Recycle to Claus plants
Toa Nenroyo, Hatsushima	October 1974	Claus plant	17	(10,000)	in 18,580 out 250	Recycle to Claus plant
Toyo Rayon, Nagoya	December 1974	Oil-fired boiler	330	(218,000)	in 1,500 out 150	Sulfuric acid
Japan National Railway Kawasaki	September 1974	Steam boiler (200 MW-3% S fuel oil)	700	(435,000)	in 1,500 out 45	Sulfuric acid
Mitsubishi Chemical, Mizushima	April 1975	Oil-fired boiler	628	(373,000)	in 1,500 out 150	Sulfuric acid
Kurashiki Rayon, Okayama	July 1975	Oil-fired boiler	410	(248,000)	in 1,500 out 150	Sulfuric acid
Fuji Film, Fujinomiya	1974	Oil-fired boiler	150	(89,000)	in 1,300 out 125	Liquid SO_2
Shin Daikyowa, Yokkaichi	December 1975	Oil-fired boiler	400	(253,000)	in 1,500 out 150	Sulfuric acid
Sumitomo Chemical, Niihama	February 1976	Oil-fired boiler	155	(91,000)	in 1,600 out 130	Sulfuric acid
Mitsubishi Chemical, Mizushima	August 1976	Oil-fired boiler	628	(390,000)	in 1,300 out 130	Sulfuric acid
Mitsubishi Chemical, Kurosaki	September 1976	Oil-fired boiler	530	(330,000)	in 1,500 out 75	Sulfuric acid
Tohoku Electric Power, Niigata	March 1977	Oil-fired boiler	380	(236,000)	in 1,000 out 100	Sulfuric acid

Table 2.2.3-3. SUMMARY OF WELLMAN-LORD SYSTEMS PLANNED IN THE U. S. [152]

System Operator and Location	Size (MWe)	New or Retrofit	Startup Mo / Yr	Disposition of SO_2
Public Service Co. of New Mexico San Juan #3	468	New	January '81	Acid
Public Service Co. of New Mexico San Juan #4	472	New	January '82	Acid
Delaware Power & Light Delaware City, 1,2,+ 3	180	Retrofit	April '80	—

nature of the source is such that the FGD system has had to follow rapid load swings from 35 to 105 percent of plant nameplate capacity, thus its flexibility and turndown capabilities are well proven. The flexibility is obtained mainly by the provision of large solution surge storage.[153]

TABLE 2.2.3-4. DISTRIBUTION OF WELLMAN-LORD SULFUR DIOXIDE REMOVAL PLANTS (Operational and Planned)[154,155]

Country	Coal-fired boilers		Oil-fired boilers		Claus/Acid plant	Total
	Industrial	Utility	Industrial	Utility		
USA	-	6	-	-	6	12
Japan	-	-	11	3	3	17
W. Germany	-	2	-	-	-	2
TOTAL	0	8	11	3	9	31

The NIPSCO plant is important because it is the first large-scale installation in the U.S. and the first full-scale demonstration of the process on a power plant burning high sulfur coal. The Wellman-Lord unit is installed on the Number 11 boiler at the Mitchell station. The nameplate rating of the boiler is 115 MW and it burned 2.9 percent sulfur coal during its demonstration.

Another installation of significance is the unit to go on-line at New Mexico Public Service's San Juan Station. This unit is designed to treat flue gas from a 314 MW boiler burning low sulfur New Mexico coal. The Allied process will be used to produce sulfur from the concentrated SO_2. Currently, New Mexico Public Service Co. is planning to install Wellman-Lord systems on an additional 1200 MW of electrical generating capacity with planned start-up date from July 1978 to January 1982.[156]

Current areas of concern are improvement in evaporator and purge treatment design. The fractional crystallization method of purge concentration was developed by Davy Powergas' two Japanese licensees. Some effort has also been spent to devise an economical means of converting the sodium sulfate to a form such as sodium carbonate or caustic soda which could be returned to the system in order to reduce the sodium makeup requirements. So far this work has not been completed. Ionics' SULFOMAT electrolytic cell and various techniques from the paper industry have shown that sodium sulfate conversion or reduction is possible, but the economics are still uncertain and some technical problems remain.

The use of an antioxidant has been evaluted by the process developer in order to reduce the quantity of sodium sulfate formed. However, the chemical compounds which were able to substantially decrease the oxidation rate turned out to be so expensive that the cost of using the antioxidants was greater than the cost of simply replacing the sodium sulfite which was converted to sodium sulfate. Therefore, use of antioxidants has been dropped in favor of various operational techniques to minimize oxidation by reducing oxygen transfer.

Data for selection of the materials of construction is plentiful due to the large amount of previous experience. The materials which will be employed are generally carbon steel lined with fiberglass reinforced polyester (FRP), 304, 316 and 316L stainless steel, all of which are fairly common. The only area where material problems are still encountered is the prescrubber. There the chlorides and ash may make the use of Hastelloy G necessary.

C. Applicability of Wellman-Lord (W-L)--Worldwide, Wellman-Lord has been applied to industrial and power boilers as well as to Claus sulfur units and acid manufacturing tail gases. These applications range from about 17 to 1100 MW (30,000 to 2,510,000 scfm).[157] All industrial boiler applications have been on oil-fired boilers in Japan.

Approximately 2/3 of U.S. industrial boilers are of 60,000 lb/hr (∿10 MW) steam capacity or less.[158] A particular concern in applying W-L to small coal-fired boilers is that these boilers are often stoker-fired, and may have no provision for over-firing. When the operator must burn undersized coal containing a high percentage of fine particles, as sometimes delivered, incomplete combustion occurs. To compensate for these factors, the boiler operators use high percentages of excess air. One FGD vendor measured ∿240% excess air in one flue gas test of a coal-fired industrial boiler. High excess air leads to excessive sulfite oxidation in the absorber liquor (Reaction 2.2.3-3). An increase in the sodium sulfate purge, of course, increases both disposal and soda ash makeup costs.

Since the W-L process is a clear liquor process, it offers a flexibility not easily achieved in slurry processes. Due to the high SO_2 loading of the scrubbing liquor the absorber and regenerator can be decoupled by storing absorption liquor such that a centralized regenerator may serve several absorber units.

D. Factors affecting performance--

1) Design and Operating Considerations - A design area of major concern in the Wellman-Lord process is sulfite oxidation. Other areas of concern are the evaporator-crystallizer, and the availability/cost of reducing agents.

The oxidation rate of the sulfite scrubbing solution is a function of several factors: impurities in the solution, recirculation rate, temperature and oxygen content of the flue gas, contact efficiency, pH, and solution strength.

It is probable that certain components of fly ash such as iron and manganese catalyze the oxidation and thus increase the total sulfate formation. Despite the upstream electrostatic precipitator and venturi prescrubber, some small amounts of fly ash will be picked up by the scrubbing liquor. The liquor is routinely filtered as it leaves the absorber to prevent build-up of these solids. Davy Powergas has noted about 25 percent increase in oxidation for oil-fired boiler applications over acid plant tail-gas applications where fly ash is not present. This information was obtained from data on Wellman-Lord units now in commercial operation. Short term tests on coal-fired boiler flue gas, however, indicated that fly ash from coal does not increase the oxidation rate over that obtained in oil-fired boiler applications. Long-term data on oxidation in applications with fly ash from coal-fired boilers has been obtained from the 115 MW demonstration unit at Northern Indiana Public Service Company's Dean H. Mitchell Station, where acceptance tests were completed September 14, 1977.[159] Long-term tests are currently under way. Unpublished results showed no significant effects of fly ash upon the overall system oxidation rate over that obtained from oil-fired operations.

The liquor recirculation rate on each stage of the absorber is governed by the need to maintain adequate flow for good hydraulic characteristics. If liquid flow across the trays is too low, poor distribution of liquor occurs and stagnant areas will result. Any recirculation above that needed for this reason merely increases the oxidation rate without increasing SO_2 removal. The reason for this is that SO_2 transfer from gas to liquid is controlled by gas film resistance while oxygen transfer is controlled by liquid film resistance. Increasing liquid recirculation decreases liquid film resistance to oxygen transfer.

The absorption temperature is fixed at the adiabatic saturation temperature of the gas, about 125°F. Oxygen content of the flue gas is determined by the amount of excess air fired. Flue gas NO_2 may be a significant contributor to oxidation but the major cause is direct oxidation of sulfite to sulfate by oxygen from the gas.

Oxygen transfer is impeded by salts in solution so the fresh scrubbing liquor is originally saturated with sodium sulfite. This also minimizes the amount of solution to be circulated. The pH range of 7 to 5.5 is controlled by the sulfite-bisulfite equilibrium. Contact efficiency is a function of tray design which is set by the level of SO_2 removal.

As a result of the oxidation problem, several approaches have been tried to reduce the amount of purge and the amount of makeup sodium carbonate. The most successful method so far has been the development of a fractional crystallization technique to increase the concentration of sodium sulfate in the purged solids to about 70 percent. This step was demonstrated in the U.S. for the first time at NIPSCO. It has been used three times previously in Japan. Further improvements in sulfate concentration are still possible.

2) Operating procedures - Operation of small coal-fired industrial boilers may deviate from large boiler operation because of several factors:

- Small companies will probably not have the luxury of long-term coal contracts for coal of consistent physical and combustion characteristics.
- Coal feed systems are more likely to be stoker type with accompanying fuel: air ratio control problems.
- Some boilers may not have overfiring capability.

These factors all call for incremental excess air over what a utility boiler would use. Air rates would also be set high to assure combustion for:

- the finest particle size coal,
- the highest HHV coal, and
- the highest fuel feed rate excursion at a given set point.

High excess air provides driving force to increase the sulfite-to-sulfate oxidation rate. As previously discussed this leads to increases soda ash makeup and solids disposal costs.

3) Performance *vs.* maintenance - The Wellman-Lord system uses a clear liquor for absorption and consequently does not have the deposition problems of lime/limestone slurry absorption. A Wellman-Lord representative estimates maintenance at 3.5 percent of fixed capital which, if accurate, is moderate.[162]

4) Fuel variations - As discussed under Operating Procedures, the boiler operator will ensure that the boiler does not smoke or develop a reducing zone around boiler tubes by resorting to increased excess air. This protects against short-term variations in fuel:air ratio, but leads to increased sulfite oxidation.

The main fuel characteristics that will affect Wellman-Lord operations are sulfur content and chlorine content. The effects of varying these fuel characteristics on this system are similar to the effects on other wet FGD systems and have been previously discussed in Section 2.2.1.1.

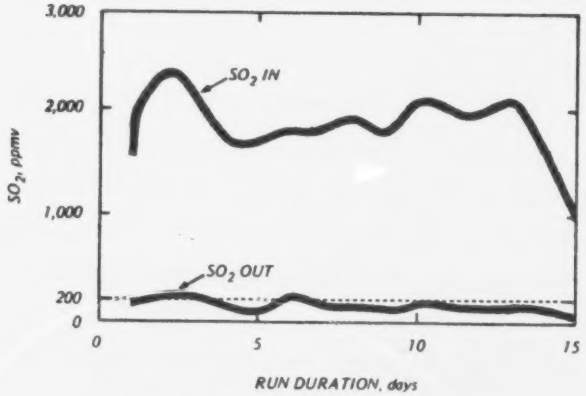
E. Retrofits--Other than problems associated with operating a Wellman-Lord system in a relatively high oxygen atmosphere, space limitations will be the major concern in retrofitting the system to industrial boilers. However, a Wellman-Lord system may be easily decoupled by adding intermediate liquor storage between the absorption and regeneration section, thus limiting the required space around the boiler. Wellman-Lord also affords the possibility of one regeneration section to economically serve several small boiler-SO_2 absorber systems if the boilers are located a reasonable distance from one another.

2.2.3.2 Wellman-Lord FGD System Performance--

A. Emission reduction--In September, 1977, the Wellman-Lord system installed at NIPSCO completed acceptance testing. Since that time, the system has been undergoing a year of extensive demonstration testing. Results of those tests have not yet been reported, but will be forthcoming in early 1979. Currently, the system is undergoing a second year of tests. Although results of the extensive testing at NIPSCO are not yet available, both Japanese and U.S. sources point to SO_2 removals of 90^+ percent and system operabilities of greater than 95 percent.[163]

Test results are, however, available from the acceptance testing conducted in 1976 and 1977. These are relatively short-term monitoring results, but they do illustrate the systems' SO_2 removal ability. In addition, EPA has recently completed a program to collect long term continous monitoring data from this system. Results of that monitoring indicated an average SO_2 removal of 89 percent throughout the monitoring period.[164]

Figures 2.2.3-2 through 2.2.3-4 illustrate three periods of sustained operations prior to the acceptance tests. As can be seen from the figures, the system effectively removed SO_2 from the flue gas for the first and second run. Poor SO_2 removals during the third run resulted from poor quality feed solution caused by mechanical problems in the soda ash feed system and evaporation area, and low feed rates to the absorber.[165]



THE SO_2 CONCENTRATION CURVES HAVE BEEN EXTRAPOLATED THROUGH DAYS 3 AND 5 BECAUSE OF INOPERATIVE INSTRUMENTATION.

Figure 2.2.3-2. Inlet and outlet SO_2 concentrations during run no. 1.[166]

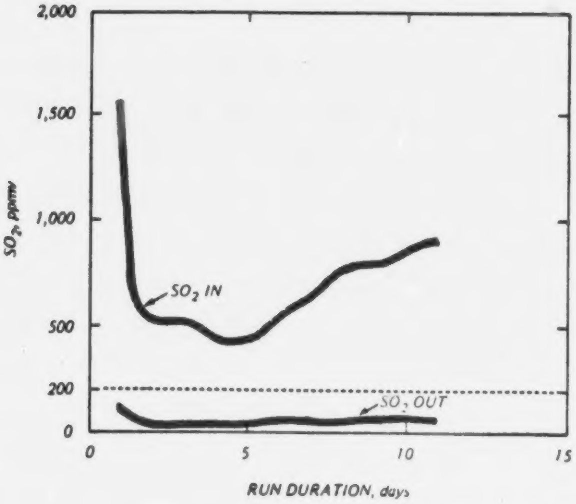
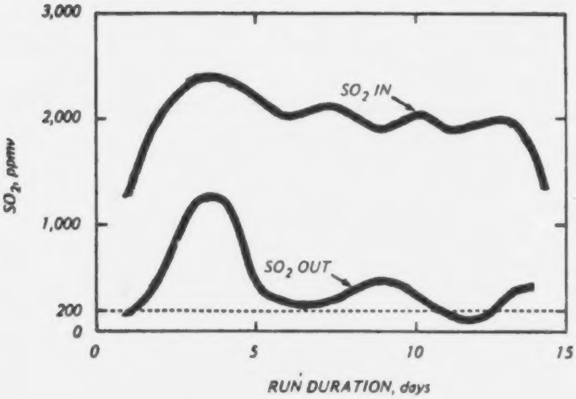


Figure 2.2.3-3. Inlet and outlet SO_2 concentrations during run no. 2.[166]



THE POOR SO_2 RECOVERIES DURING THIS PERIOD RESULTED FROM POOR QUALITY SOLUTION CAUSED BY MECHANICAL PROBLEMS IN THE SODA ASH FEED SYSTEM AND EVAPORATION AREA, AND LOW FEED RATES TO THE ABSORBER WHILE BALANCING TANK INVENTORIES.

Figure 2.2.3-4. Inlet and outlet SO_2 concentrations during run no. 3.[167]

The actual acceptance tests were divided into two phases. Phase I was a 12 day test at an average flue gas flow of 320,000 acfm, and Phase II was an 83 hour test at an average flue gas flow of 390,000 acfm. Results of the Phase I tests are shown in Figure 2.2.3-5. This figure shows that the SO_2 removal of the system was in excess of 90 percent (the minimum acceptable level) throughout the test period. In addition, particulate emissions from the absorber remained below the Federal NSPS of 0.1 lb/10^6 Btu throughout the test period.

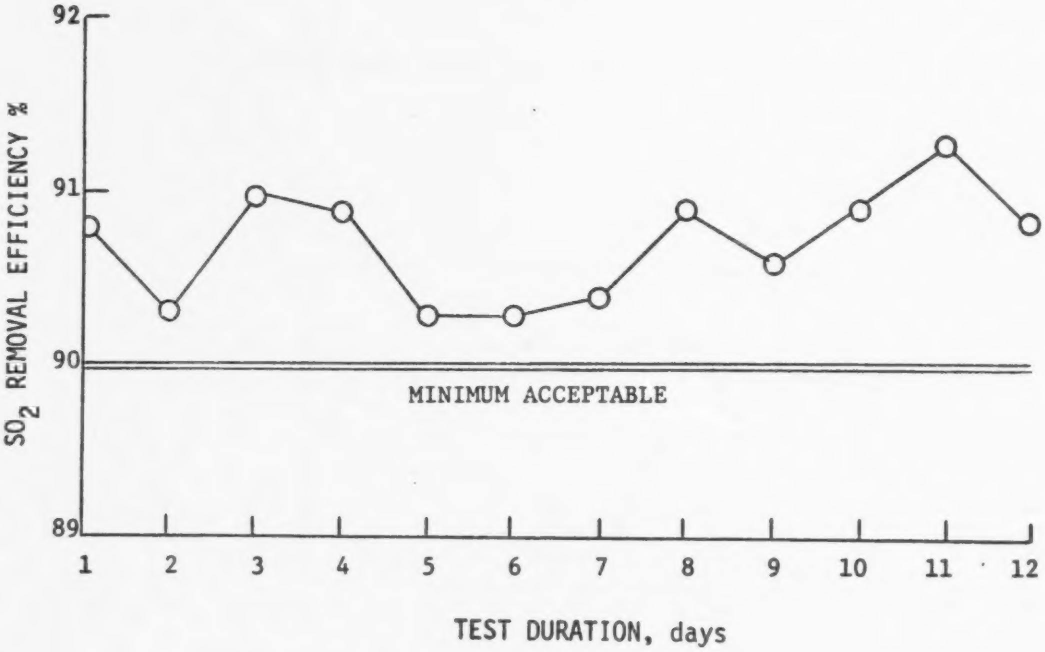


Figure 2.2.3-5. SO_2 removal efficiencies during 12-day test.[168]

During the Phase II, 83 hour test, the FGD system was also required to achieve an SO_2 removal of 90 percent. Under the more stringent Phase II operating conditions, the SO_2 removal efficiency averaged 91 percent. Particulate emissions during these tests were also below the Federal NSPS limit.[169]

B. Impact of Wellman-Lord performance on boiler performance--Major effects of a Wellman-Lord system on a boiler's performance would be: 1) boiler derating due to the parasitic energy required to run the pumps and fans and to regenerate the sodium sulfite sorbent, and 2) boiler load reductions due to Wellman-Lord downtimes, assuming no bypass of flue gas is allowed. Energy consumption for a 500 MW rated W-L system was reported to be 32 MW, not including SO_2 conversion.[170] This amounts to a derating of 6.4 percent. Calculations were performed to estimate the energy requirements for industrial boiler application (see Chapter 5) which showed the process to consume from 3 to 8 percent of the net heat input to the boiler, depending mainly upon the amount of SO_2 removal.

C. Additional maintenance requirements--Wellman estimates maintenance cost at 3.5% of fixed capital.[171] Since the W-L process is a clear liquor process, it should not be expected to have excessive maintenance problems in the absorber section. It is, however, more mechanically complex than a once-through system such as lime/limestone slurry FGD.

2.2.4 Magnesia Slurry Absorption Process

2.2.4.1 System Description--

The Magnesia Slurry Absorption Process uses magnesium hydroxide to absorb SO_2 in a wet scrubber. Magnesium sulfite is the predominant species formed by the reaction with SO_2 in the scrubber according to Equation 2.2.4-1.

$$Mg(OH)_2 + SO_2 \rightarrow MgSO_3 + H_2O \qquad (2.2.4\text{-}1)$$

Reaction 2.2.4-2 occurs to a lesser extent.

$$MgSO_3 + H_2O + SO_2 \rightarrow Mg(HSO_3)_2 \qquad (2.2.4\text{-}2)$$

The aqueous slurry is centrifuged and the cake is then dried with fuel oil combustion gas to remove free and bound moisture. The magnesium oxide is regenerated in a calciner by thermal decomposition of magnesium sulfite according to Equation 2.2.4-3.

$$MgSO_3 \rightarrow MgO + SO_2 \qquad (2.2.4\text{-}3)$$

Hot combustion gases are again used to heat the magnesium sulfite in the calciner and to remove the SO_2. The SO_2 gas stream may be used to produce sulfuric acid or elemental sulfur.

Three magnesia-based wet scrubbing processes have been developed since the early 1930's: 1) A basic (high pH) $MgSO_3/Mg(OH)_2$ slurry process, 2) an MnO_2 activated absorbent slurry system, and 3) an $MgSO_3$ acidic clear liquor process. The basic slurry process is the most advanced system and, therefore, will be discussed in this chapter.

A. Process description--The basic magnesia scrubbing process can be divided into four major process areas: SO_2 absorption, $MgSO_3/MgSO_4$ separation and drying, MgO regeneration and SO_2 recovery, and SO_2 conversion. Figure 2.2.4-1 is a simplified flow diagram for the process.[172]

1) SO_2 absorption - Absorption of SO_2 takes place after the flue gas is treated for particulate removal in a wet scrubber or electrostatic precipitator. A separate system ahead of the SO_2 scrubber is used for particulate removal for two reasons. First, it eliminates some components of the fly ash such as vanadium and iron compounds which can catalyze the oxidation of $MgSO_3$ to $MgSO_4$. Also, there is no easy way to remove fly ash from the circulating scrubbing slurry.

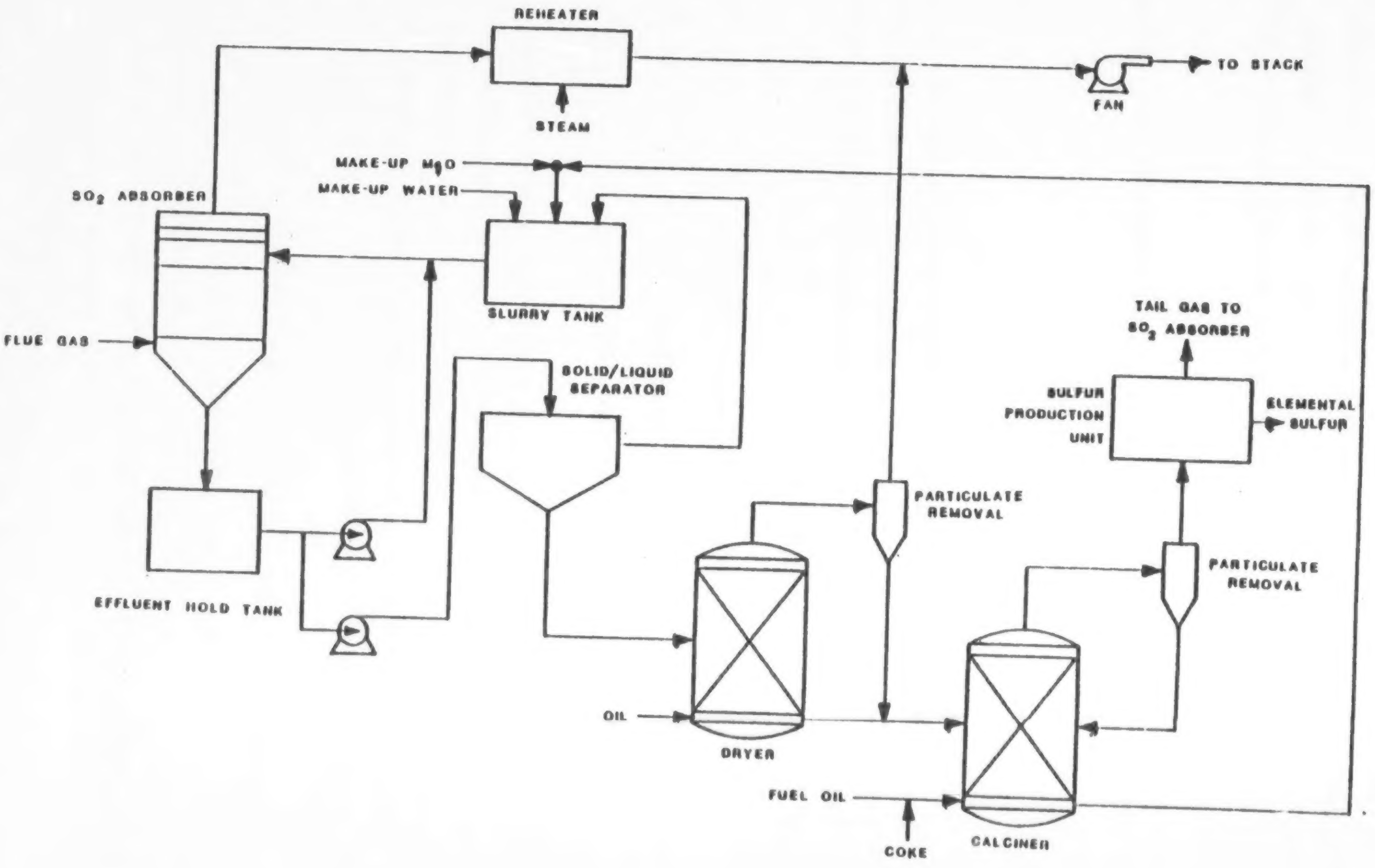


Figure 2.2.4-1. Process Flow Diagram for the Magnesia Slurry Absorption Process.

An aqueous slurry of magnesium hydroxide and magnesium sulfite (pH range 6.5 to 8.5) is used to absorb the SO_2 according to Equations 2.2.4-1 and 2.2.4-2. The $MgSO_3$ is formed as a crystalline solid in slurry. A bleed stream is sent to a centrifuge as a first step for MgO recovery. Makeup water, recycle MgO, and makeup MgO are added to the recycle slurry to maintain a constant slurry composition.[173]

Sulfite oxidation gives rise to sulfates in the system by the following reaction:

$$MgSO_3 + \tfrac{1}{2}O_2 \rightarrow MgSO_4 \qquad (2.2.4\text{-}4)$$

The test facility at PEPCO-Dickerson reported sulfate concentrations to be about 7 mole percent for the anhydrous solids transferred to the regeneration section.[174] The sulfite and sulfate solids precipitate as hydrated crystals as illustrated in the following equations:

$$MgSO_3 + 6H_2O \rightarrow MgSO_3 \cdot 6H_2O \qquad (2.2.4\text{-}5)$$

$$MgSO_3 + 3H_2O \rightarrow MgSO_3 \cdot 3H_2O \qquad (2.2.4\text{-}6)$$

$$MgSO_4 + 7H_2O \rightarrow MgSO_4 \cdot 7H_2O \qquad (2.2.4\text{-}7)$$

$MgSO_3 \cdot 6H_2O$ is the preferred form because of its large, easily separable crystalline form. Laboratory work indicates that it can be preferentially formed, given proper design and operating conditions.[175] The bisulfite in the spent scrubbing liquor is reacted with magnesium hydroxide which is formed by slaking the fresh and recycle magnesium oxide.

$$Mg(HSO_3)_2 + Mg(OH)_2 + 4H_2O \rightarrow 2(MgSO_3 \cdot 3H_2O) \qquad (2.2.4\text{-}8)$$

$$MgO + H_2O \rightarrow Mg(OH)_2 \qquad (2.2.4\text{-}9)$$

2) $MgSO_3/MgSO_4$ separation and drying - After absorption of SO_2 in the scrubber, a portion of the slurry from the main scrubber circulation loop is

removed as a 10-15 weight percent slurry and sent to a thickener. The thickener is an optional piece of equipment that has the potential for increasing the surge capacity in the recovery portion of the plant and for improving centrifuge operation.[176]

A stainless steel, solid bowl centrifuge has been used to recover a wet cake of $MgSO_3/MgSO_4$ hydrate crystals. Satisfactory centrifuge operation is necessary so that solids are removed at a sufficient rate to maintain control of recycle solids concentration. Wet crystals are discharged from the centrifuge through a vertical chute into a screw feeder which provides a seal and a continuous flow of wet solids into a rotary fluid-bed dryer. The rotary kiln type dryer is presently used in the three U.S. magnesia scrubbing demonstration units. Combustion gas from an oil burner, which can be tempered by a sidestream of stack gas, is used to dry the crystals. The dried $MgSO_3/MgSO_4$ is discharged from the dryer and conveyed to a calciner for Mgo regeneration and SO_2 recovery.

3) MgO regeneration and SO_2 recovery - Dried $MgSO_3/MgSO_4$ solids are heated in an oil-fired rotary kiln or fluidized-bed reactor until decomposed. The main decomposition reaction is shown in Equation 2.2.4-3. The $MgSO_4$ is also reduced in the calciner using carbon as a reducing agent.

$$MgSO_4 + \tfrac{1}{2}C \rightarrow MgO + SO_2 + \tfrac{1}{2}CO_2 \qquad (2.2.4\text{-}10)$$

Two installations have used a rotary kiln to regenerate the MgO. High dust losses in the rotary kiln require the use of a hot cyclone and venturi scrubber to recover all of the MgO. If a fluidized-bed reactor were used most of the MgO formed would go overhead with the SO_2 and combustion gases and separation equipment would also be required.

The optimum calcining temperature in the reactor is set by the fact that it must be high enough to decompose all of the $MgSO_3/MgSO_4$ solids without "dead burning" the MgO. "Dead burned" MgO is that which has been melted into a refractory like material and is chemically unreactive and not effective for further SO_2 removal. Operating temperatures in the 815°C (1500°F) range have been used for rotary calciners in this service.

4) SO₂ conversion - After dust removal, the sulfur dioxide rich gas from the calciner is piped to either a sulfur or sulfuric acid production unit. This concentrated SO_2 stream is actually not as well suited for sulfur production as it is for acid production. This is because of the oxygen in the concentrated SO_2, which is introduced by the excess air used for combustion of fuel oil in the calciner. This oxygen will consume additional reductant than what is required for reduction of SO_2 to sulfur, thus increasing sulfur production costs. However, the SO_2 stream is well suited for acid production, as it is at approximately 38°C (100°F), is saturated with water, and contains 8-10 percent SO_2.

B. Status of development--The magnesia slurry scrubbing process has been shown to be feasible on a full-scale size. Three retrofit units in the U.S. of the 95-150 MW size have demonstrated greater than 90 percent SO_2 removal on both oil-fired and coal-fired systems.[177,178] A list of the operating and planned magnesia scrubbing units is shown in Table 2.2.4-1. MgO units make up about 5 percent of total U.S. FGD capacity. None of the U.S. applications of MgO technology are on industrial boilers.[179]

The magnesia scrubbing process has been used on a commercial scale at three locations in Japan, and constitutes about 1 percent of total FGD capacity.[180] A summary of the status of the three Japanese installations is shown in Table 2.2.4-2. SO_2 removals of 90 to 99 percent have been demonstrated for the Japanese units. The Mitsui unit uses a special cross-flow type absorber which they have developed. The scrubber consists of an empty chamber with two rotating shafts with many spoons. The scrubber has little

TABLE 2.2.4-1. OPERATING AND PLANNED MAGNESIA SCRUBBING UNITS ON U.S. POWER PLANTS AS OF AUGUST 1978[181, 182, 183]

Utility Company, Power Station	New or Retrofit	Size of FGD Unit (MW)	Process Vendor	Fuel, Sulfur Content, (%)	SO_2 Recovery (%)	Start-Up Date, Status
Boston Edison, Mystic No. 6	Retrofit	150	Chemico	Oil, 2.5	90*	Start-up in April, 1972; Test program completed in June, 1974. Not currently operational.
Potomac Electric and Power, Dickerson No. 3	Retrofit	95	Chemico	Coal, 2.0	90†	Start-up in September, 1973; Test program completed in September, 1975. Not currently operational.
Philadelphia Electric Co. Eddystone No. 1A	Retrofit	120	United Engineers	Coal, 2.5	96-99	Start-up in September, 1975; Currently operational.
Philadelphia Electric Co., Eddystone No. 1B	Retrofit	240	United Engineers	Coal, 2.5		Start-up due in June, 1980; Letter of intent signed.
Philadelphia Electric Co., Eddystone No. 2	Retrofit	336	United Engineers	Coal, 2.4		Start-up due in June, 1980; Considering FGD system.
Philadelphia Electric Co., Cromby	Retrofit	150	United Engineers			Start-up due in June, 1980; Considering FGD system.

* Reference 181 given long term SO_2 removal of 77-87%.

† Reference 183 gives long term SO_2 removal of 77-83%.

TABLE 2.2.4-2. OPERATING MAGNESIA SCRUBBING UNITS ON JAPANESE POWER PLANTS AS OF AUGUST 1978[184]

Company Location	Type of Plant	Size of FGD Unit (MW)	Process Vendor	Flue Gas SO_2 Content (ppm)	SO_2 Recovery (%)	By-Product (tons/day)	Start-Up Date, Status
Onahama Smelter Onahama, Japan	Copper Smelter	28	Onahama-Tsukishima	15,000-25,000	99.5	H_2SO_4-240	Start-up in December, 1972; Currently operational.
Mitsui Mining Hibi, Japan	H_2SO_4	24	Mitsui-Mining	1,500-2,000	90	H_2SO_4-18	Start-up in October, 1971; Currently operational.
Idemitsu Kosan Chiba, Japan	Claus Unit and Boiler	162	Chemico-Mitsui	---	95	Sulfur-70	Start-up in 1974; Currently operational.

possibility of scaling because of the simple structure, but the size of the scrubber is limited to the treatment of no more than 1700 Nm /min (60,000 scfm) or about 29 MW.

The magnesium oxide FGD system at Boston Edison has been able to demonstrate 80 percent availability during sustained operation. The Potomac Electric Dickerson station was only able to operate at 64 percent availability for their SO_2 recovery system during their best month.

C. Applicability of the magnesia slurry FGD process (MgO)--Magnesia slurry has been used in Japan for oil-fired power plant flue gas cleanup and on industrial tail gases.[185] In the U.S. the process has been applied to both oil- and coal-fired utility boilers but has not been used on an industrial boiler.[186] The relative complexity of this process may limit application to larger industrial boilers or to multiple-boiler situations where multiple absorbers and a common regenerator might be installed.

D. Factors affecting performance--

1) Design and operating considerations - It is not entirely clear which design factors have most influenced the contrasting performance of magnesia FGD systems as reported. Generally, high SO_2 removals are achievable because magnesium sulfite is relatively soluble. The amount of soluble alkalinity available in magnesia scrubbing systems is more than that available in lime/limestone systems but less than for sodium-based scrubbing systems. Therefore, magnesia systems can operate at a lower L/G than lime/limestone systems but at a higher L/G than sodium-based systems. An L/G of 33 gal/1000 acf was used at the Boston Edison facility using a venturi scrubber.[187]

The major design variables that can be used to regulate the scrubbing operation are the recycle pH and percent solids. A higher pH gives a higher SO_2 recovery. The pH can be controlled by MgO addition while the amount of solids in the slurry can be controlled by adjusting the rate of bleed from the recycle slurry.

The process sequence for magnesia slurry scrubbing has been fairly standard for the U.S. installations that have already been operated. A centrifuge and rotary dryer have been used in all three U.S. demonstration plants. Two of the installations used a rotary calciner while Philadelphia Electric uses a fluidized bed unit to calcine the $MgSO_3/MgSO_4$ product. A thickener is an optional piece of equipment that has been used to concentrate the slurry prior to the centrifuge. The use of a thickener gives the prospect for improved centrifuge operation.

Particulate control is carried out separately from the SO_2 scrubbing system. High dust removal efficiencies are required to minimize the amount of contaminants that enter the system. Small amounts of fly ash will still enter the circulating slurry system since particulate control devices are not 100 percent effective in removing fly ash.

Contaminant control is required to reduce the amount of impurities that will accumulate in closed-loop operation. Both soluble and insoluble impurities are of concern. Insoluble impurities come primarily from the fly ash in the flue gas and coke used in the calciner. These insoluble impurities are controlled by high efficiency particulate removal equipment upstream of the SO_2 absorber and by the use of coke with a low ash content.

Soluble contaminants enter the system from the makeup water, makeup MgO, fly ash, and calciner coke. These soluble species must be purged from the system. Chloride attack of scrubber internals can occur when carbon steel or stainless steel are used as absorber construction materials; consequently a prescrubber can be used for chloride removal. Glass reinforced polyester resins have been successfully used to prevent corrosion in the scrubber.

Heat recovery from the dryer and calciner off gas is desirable but is complicated by the entrainment of $MgSO_3$ or MgO fines in the combustion gases. The dryer off gas at 200°C (400°F) and the calciner off gas at 870°C (1600°F) offer a useful source of energy savings for the process.

2) Operating procedures - Numerous operating problems occurred early in the program at Boston Edison. Most of the problems were of a materials handling nature resulting from the characteristics of the solids generated in the scrubber loop. It was found that trihydrate crystals with an average size of 10-15 microns were formed in the absorber instead of the larger hexahydrate which had been formed in the pilot plant. As a result, the centrifuge cake contained as much as 25 percent unbound moisture. This led to problems of solids adhesion to the dryer drum.

These problems were solved by several operating and design modifications. Among these modifications were changing the dryer to function as a granulator and adding hammers to loosen any material which tended to adhere to the dryer shell. The granulator material discharge was screened and sent through lump breakers to eliminate oversize agglomerated granules of the magnesium sulfite. The dryer off gas was sent to the SO_2 absorber to prevent high dust losses. This subsequently caused the loss of 8°C (15°F) of reheat of the saturated flue gas expected from the dryer off gas.

Other process problems occurred in the calcining system. A rotary kiln has been used for these operations. The formation of the very fine trihydrate crystals in the oil-fired power plant application also resulted in dusting problems in the rotary calciner. The facility at Essex Chemical resolved the dusting problem in the calciner by the use of a cyclone followed by a venturi scrubber to remove all of the MgO fines from the gas. Leakage of air into the calciner was a problem since high oxygen levels interfere with the reduction of $MgSO_4$. Installation of new seals on the rotary calciner corrected the problem.

In addition, high calciner temperatures can cause sintering or "dead burning" of the regenerated MgO which will result in unreactivity of the product for reuse in the scrubber. High reactivity magnesia is favored by low calciner temperature and by increased amounts of carbon in the calciner feed. Recycle MgO reactivity was improved by the correct calciner operating conditions, by pulverizing the calcined MgO, and by heating the MgO slurry tank. More recent tests indicate that continuous stable operations are possible with high SO_2 removal efficiencies and that continuous use of recycle MgO has only a slight effect on the system.

The chemical and mechanical performance of the scrubber was excellent at Boston Edison. No internal plugging was encountered and the polyester lining of the scrubber was in sound condition after two years of intermittent operation.[188]

Erosion and corrosion were experienced in the carbon steel recirculating slurry piping. The use of rubber-lined pumps, valves, and piping in certain areas of the system is considered to be a practical solution to this problem. The slurry recirculating pumps in the system have withstood corrosion using 316 stainless steel impellers.[189]

Potomac Electric's coal-fired Dickerson Station FGD system achieved only 78-83 percent SO_2 removal and 48 percent availability at 75 percent design over a five month test period.[190] The maximum Dickerson stream time without shutdown from equipment malfunction was 120 hours.[191] Although corrosion, erosion, leaks, centrifuge outages, etc. were mentioned as contributing to low reliability, no figures were given to indicate outage time as a function of the type of equipment failure, except for leaks. A report of Dickerson operations revealed during the test period there were nine major outages, three of which were leaks, and that there were forty-seven leaks observed during four months of record keeping, sixteen of which induced shutdowns. Plant construction with off-spec pipe, fittings, and rubber linings was blamed for many equipment failures.[192]

3) Performance vs. maintenance - Maintenance problems at the coal-fired Dickerson plant appear to have stemmed primarily from erosion/corrosion These problems were in part blamed on construction with off-specification pipe, fittings and insufficient rubber piping liners. Over 40 leaks occurred in approximately 5 months of testing.

4) Fuel variations - The long-term problem of acquiring coal of consistent quality applies to owners of all small-scale industrial boilers, who will probably not be able to write the same long-term contract that a large utility company might. The day-to-day impact is that higher excess air may be introduced to the boiler than would be used in a utility boiler. As a result, the flue gas will have a lower SO_2 concentration, and the absorber will be larger relative to the amount of SO_2 to be removed, for a given heat release, coal sulfur content, and SO_2 removal efficiency.

2.2.4.2 Magnesia Slurry FGD System Performance--

A. Emission reductions--Only limited data are available from the full-scale systems that show the interrelationship of SO_2 removal to the various process variables. Operation of these units did, however, show that 90 percent control of SO_2 was achievable. Part of the reason that data are not readily available from these systems is the relatively low system availabilities obtained at these installations.

For a given L/G the effect of inlet SO_2 concentration on SO_2 removal efficiency was relatively minimal at high pressure drops at the 155 MW Mystic facility, as shown in Figure 2.2.4-2. At lower pressure drops across the venturi absorber significant effects were noted over the 400-2000 ppm SO_2 range. The relationship between SO_2 removal efficiency and pressure drop is more directly illustrated in Figure 2.2.4-3.

In 1970 Babcock and Wilcox evaluated SO_2 absorption responses to various process variables using both a floating bed absorber and a venturi scrubber/absorber. Devitt, *et al.*, report that in order to achieve a 90 percent SO_2 removal efficiency, the pH of the scrubbing slurry should be maintained

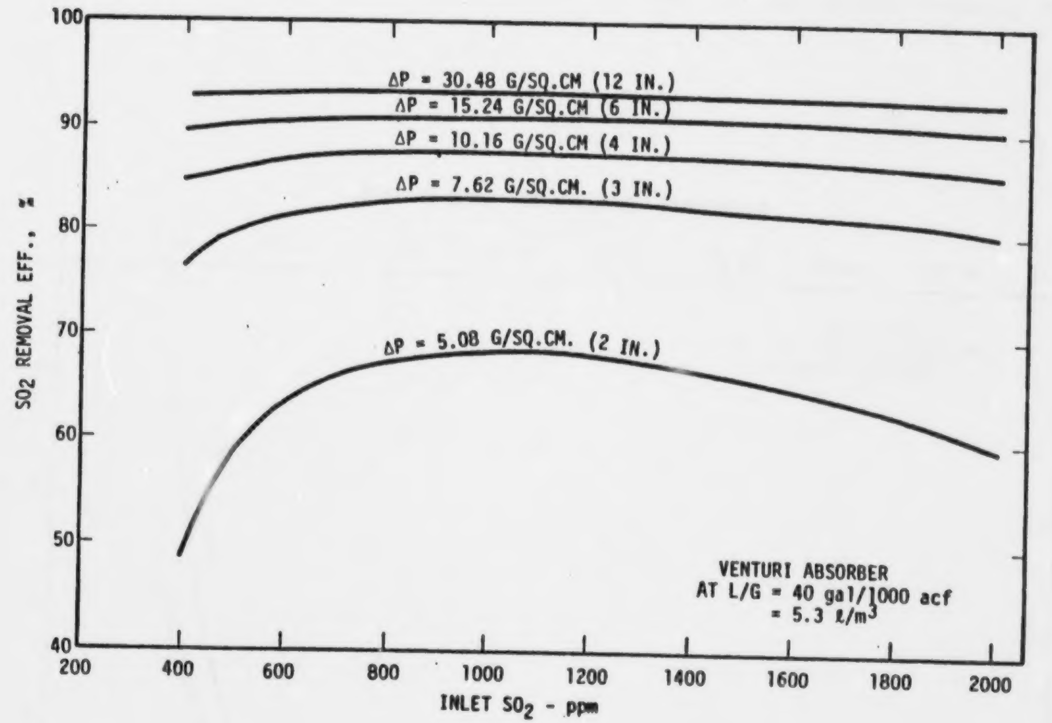


Figure 2.2.4-2. Effect of inlet SO_2 concentration and venturi pressure drop on SO_2 removal for the Mystic venturi absorber.[197]

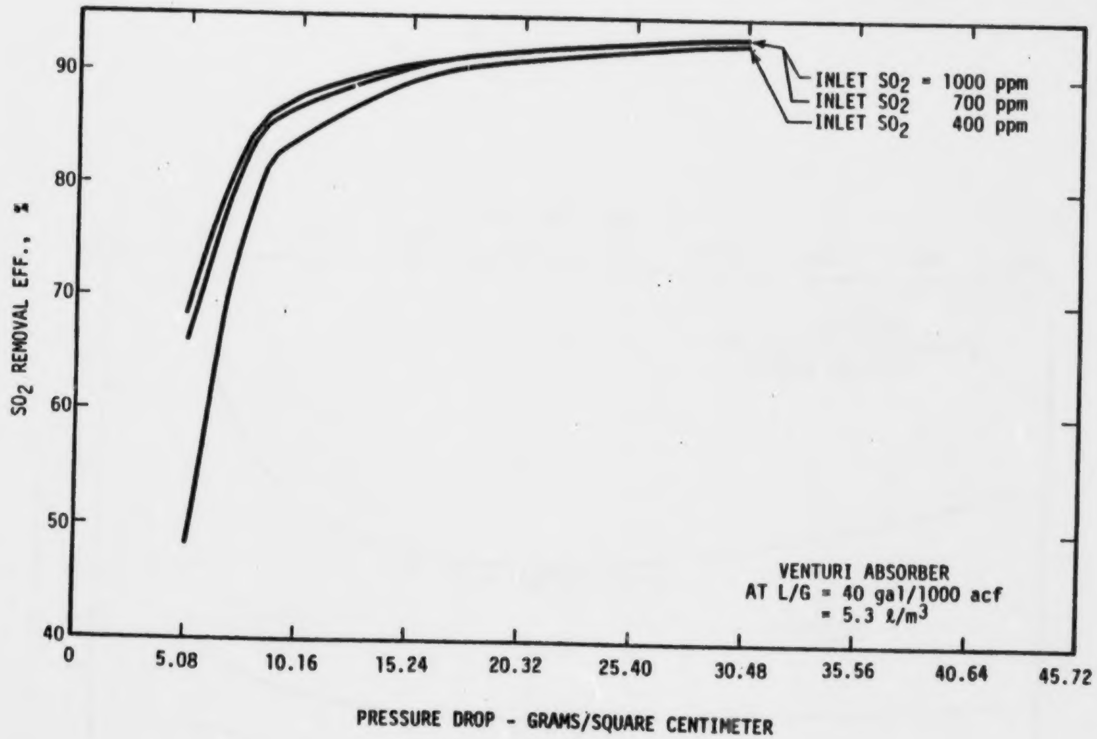


Figure 2.2.4-3. Effect of pressure drop on SO_2 removal for the Mystic venturi absorber.[198]

between 6.0 and 7.5.[195] This is based on data reported by Semrau shown in Figure 2.2.4-4. A 227 kg/hr (500 lb/hr) coal feed rate was used for these tests. Results showed that SO_2 absorption efficiency was a function of the bisulfite concentration and pH of the recycle liquid for a given absorber L/G. Additional work, which was not particularly successful, attempted to rationalize the mass transfer coefficient determined from this work to fundamentally developed values.[196]

B. Impacts on boiler--The major impacts of a magnesium oxide FGD system on boiler operations are similar to those of other wet FGD systems; i.e., 1) power consumption for running the system's pumps and fans, and 2) possible boiler load reduction during FGD system outages if no bypass is permitted. The power consumption for a 500 MW magnesium oxide unit was reported to be about 1.8 percent of the unit generating capacity. This consumption was for the power to operate the systems pumps and fans and excluded other energy requirements such as fuel oil for drying and calcining the magnesium sulfite solids.

Boiler derating due to load reductions, however, is more difficult to evaluate based on U.S. operating experience. Past U.S. experiences with magnesium oxide systems indicate that this impact may be more significant than for other FGD systems; however, Japanese magnesium oxide systems have operated with high reliabilities.

C. Additional maintenance requirements--Reported estimates of maintenance costs for the magnesium oxide process vary from four to seven percent of fixed capital.[200] This is higher than maintenance costs reported for Wellman-Lord (3.5 percent). Experience to date in the U.S. points to problems of corrosion and erosion of metal parts as the most clearly defined maintenance problem affecting system performance.

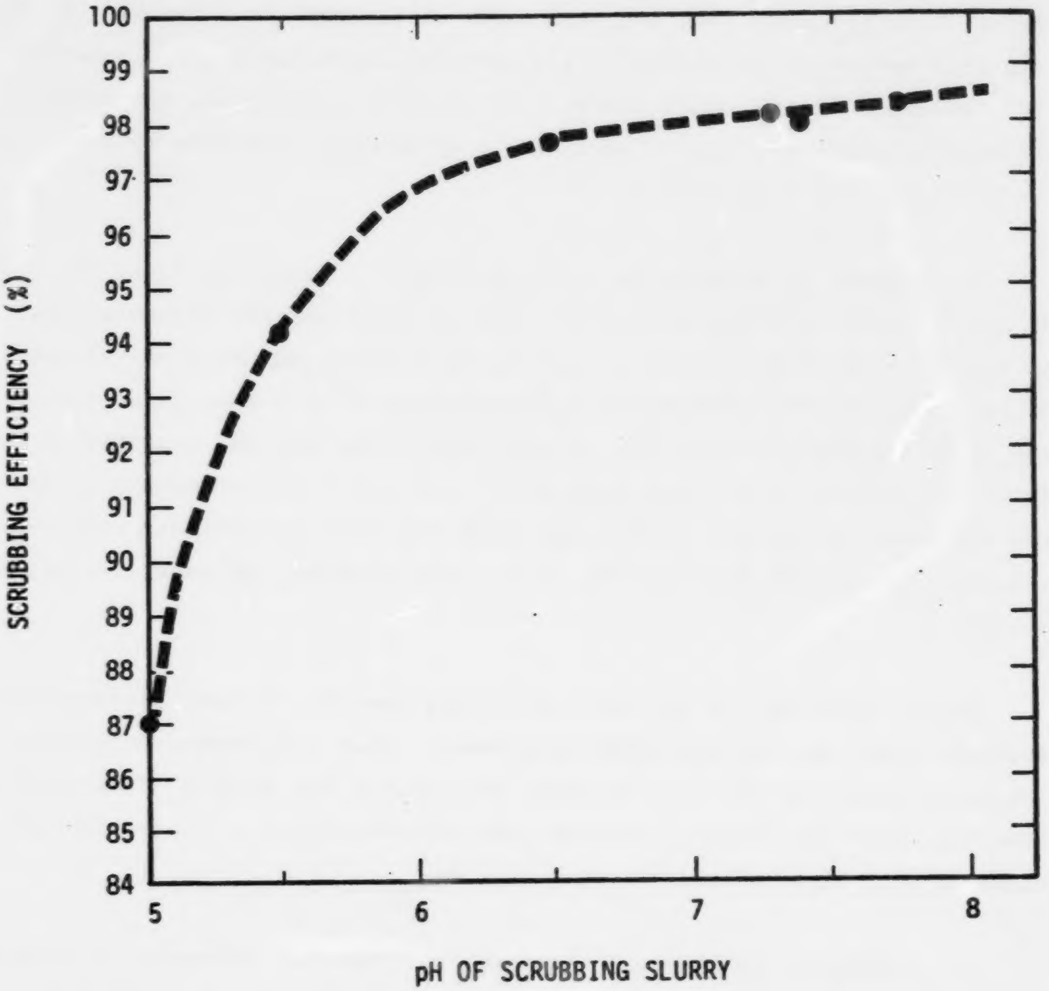


Figure 2.2.4-4. The effect of pH on SO_2 scrubbing efficiency.[199]

2.2.5 Sodium Scrubbing

The sodium scrubbing process is capable of achieving high SO_2 removal efficiencies over a wide range of inlet SO_2 concentrations. The process consumes a premium chemical and produces a soluble waste salt which under current practice, is normally discharged to a lined evaporation pond for drying.

2.2.5.1 System Description--

A. System--Sodium scrubbing processes currently being used for flue gas desulfurization (FGD) on utility and industrial boilers employ a wet scrubbing solution of NaOH, Na_2CO_3 or $NaHCO_3$ to absorb SO_2 from the flue gas. The operation of the wet scrubber is characterized by a low liquid-to-gas ratio (L/G) and a clear scrubbing liquid because of the high solubility of sodium salts. The absorption reactions which take place in the scrubber are:

$$2NaOH(\ell) + SO_2(g) \rightarrow Na_2SO_3(\ell) + H\ ) \qquad (2.2.5\text{-}1)$$

$$Na_2CO_3(\ell) + SO_2(g) \rightarrow Na_2SO_3(\ell) + CO_2(g) \qquad (2.2.5\text{-}2)$$

$$2NaHCO_2 + SO_2(g) \rightarrow Na_2SO_3(\ell) + H_2O + 2CO_2(g) \qquad (2.2.5\text{-}3)$$

$$Na_2SO_3 + SO_2(g) + H_2O \rightarrow 2NaHSO_3 \qquad (2.2.5\text{-}4)$$

Simultaneously some sodium sulfite reacts with the oxygen in the flug gas to produce sodium sulfate:

$$Na_2SO_3(\ell) + \tfrac{1}{2}O_2(g) \rightarrow Na_2SO_4(\ell) \qquad (2.2.5\text{-}5)$$

The scrubber effluent solution thus consists of a mixture of Na_2SO_3, $NaHSO_3$, and Na_2SO_4. Scrubber effluent from this nonregenerable process is then treated in a variety of ways. If fresh caustic or soda ash is added for makeup the majority of the effluent is recycled to the scrubber with a slipstream going to wastewater treatment and disposal. If a process waste stream is used for scrubbing, the effluent is used on a once-through basis and then disposed. Pulping operations can make use of the sulfite-bisulfite solution in their digesters. Figure 2.2.5-1 presents a simplified process diagram for a sodium scrubbing system.

Sodium scrubbing encompasses two categories; systems that use chemical addition for reagent makeup and systems that use a process waste stream for scrubbing. Chemical addition can be differentiated according to reagent, either Na_2CO_3 or NaOH. Some FGD systems that are located at a plant which produces an alkaline waste stream use the waste stream in their scrubbing process. Soda ash plants which use end liquor and pulping operations which produce a caustic waste stream are examples. Table 2.2.5-1 gives a summary of the capacities of sodium systems that are operational or under construction in the United States.

TABLE 2.2.5-1. SUMMARY OF SODIUM FGD PROCESSES
ON U.S. COAL-FIRED INDUSTRIAL AND UTILITY BOILERS[201]

System Category	Industrial Boilers			Utility Boilers	
System Category	No. Units	SCFM	MWe	No. Units	MWe
Operational	102	4,145,000	2073	3	375
Construction	19	656,000	328	1	509
Planning	4	286,000	143	-	-
Totals	125	5,087,000	2544	4	884

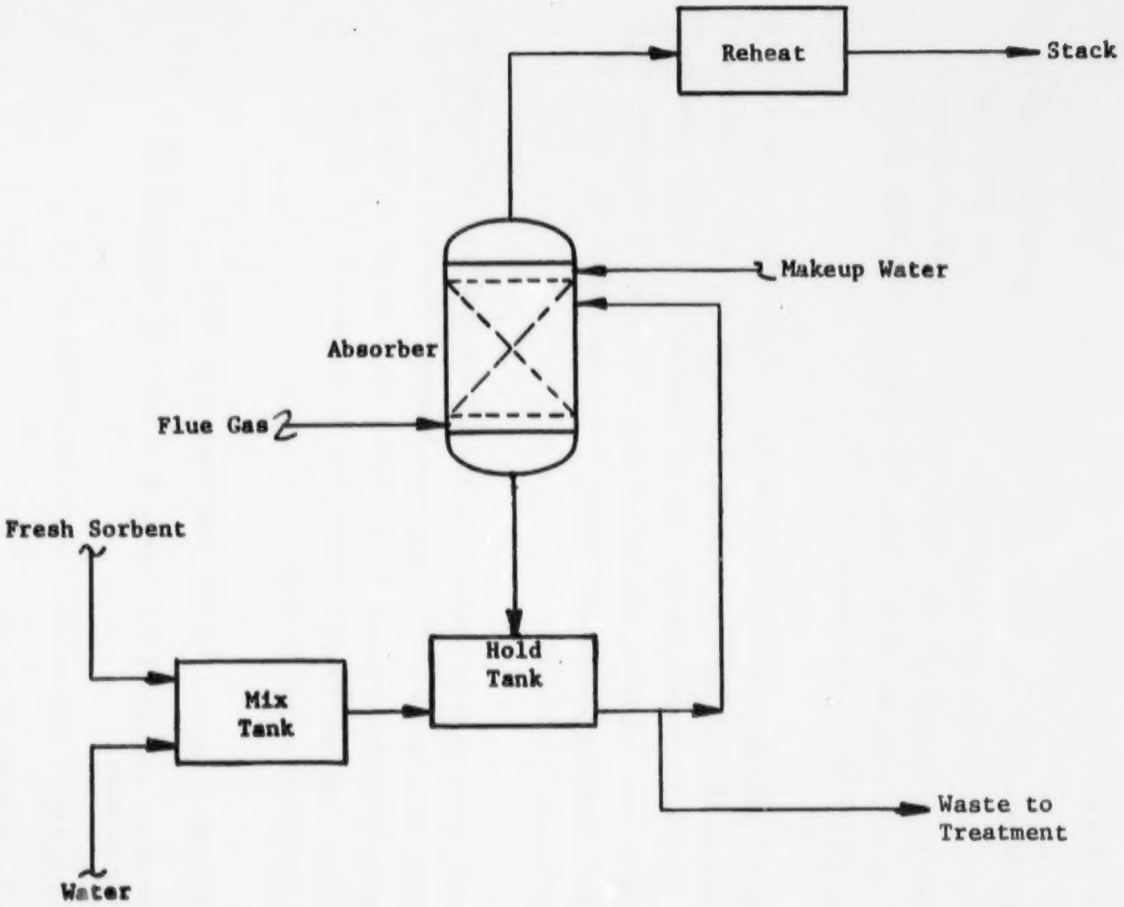


Figure 2.2.5-1. Simplified Flow Diagram Sodium Scrubbing System.

Even though the sodium scrubbing process is one of the least complex FGD processes both chemically and mechanically, some accessory equipment is required. Particulate control may or may not be used upstream of the absorber depending on the absorber type. Flue gas reheat may be necessary, and facilities for waste disposal are required.

If fly ash is not removed prior to SO_2 removal, a venturi scrubber may be used for both particulate and SO_2 removal. The FGD system must then be designed for slurry handling and must deal with special disposal considerations. If fly ash is removed prior to SO_2 removal, the absorber can be either a packed tower or a tray tower.

Solid reagent handling requirements may vary from site to site, but some generalities can be made. Storage, usually silos, is required. Transport from the silo to a mixing tank can be done by conveyor. A mixing tank is needed to dissolve the reagent. The solution is sometimes then pumped to a clarifier where insoluble impurities settle out. The clarified liquor is then ready for use in the scrubber.

Wastes from wet sodium processes contain sodium sulfite, sodium sulfate, sodium carbonate, sodium hydroxide, and some inerts. Disposal of this material presents problems because it is highly water soluble. Several disposal possibilities have been studied, but more research is required. The disposal options studied include:

- Developing a market for the resultant salt cake.
- Short term storage of the waste with eventual recovery of sodium products.
- Insolubilizing the waste by forming complex salts.
- Returning waste to mines.
- Permanent storage of waste in lined and covered pits.

At present, the trend for disposal practices is toward wastewater treatment and holding ponds for evaporation. The specifics of the wastewater treatment from users are not reported.[202] However, a general scheme for the treatment can be outlined from several sources. The liquor must be clarified to remove fly ash and solids. The clarified liquor may then be aerated to oxidize sulfite to sulfate to reduce chemical oxygen demand. Solution pH is then adjusted and the liquor discharged to sewers or rivers.

B. Development status--Sodium scrubbing systems are considered commercialized technology. These systems have been developed in both the U.S. and Japan. The first full-scale application of sodium FGD in the U.S. was at a General Motors plant in St. Louis, Missouri. Two systems were installed on coal-fired boilers in 1972. In the U.S. sodium carbonate scrubbing is commercially practiced on 375 MW of generating power at the Nevada Power Reid Gardner Station. The Reid Gardner Station has operated 250 MW of the system since April 1974 and 125 MW of the system since July 1976. The system has operated well without any major operating problems. FMC's soda ash plant in Green River, Wyoming has had two scrubbers on two 100 MW (equivalent) units since May 1976. Table 2.2.5-2 gives a summary of FGD processes on U.S. coal-fired utility boilers. This table shows that in the utility sector, sodium carbonate scrubbing is used on 375 MW from a possible 14,420 MW of operational systems (about 3 percent). However, Table 2.2.5-3 shows that on industrial boilers, sodium scrubbing is the preferred FGD process. Use of sodium scrubbing on industrial boilers is because of several attractive features of the system:

- It can use a process waste stream as the SO_2 sorbent.
- It uses a clear liquor rather than slurry which lessens the potential for plugging and scaling.
- It is more tolerant of changes in boiler load conditions.

Many of the sodium systems on industrial boilers were the result of adding NaOH to the circulating liquor in particular scrubbers for pH control. It was found that adding NaOH for pH control not only aided in minimizing corrosion, it also increased SO_2 removal to the point that the particulate scrubber assumed a dual role of both particulate and SO_2 removal. Table 2.2.5-4 presents a summary of operating sodium scrubbing systems applied to U.S. industrial boilers.

TABLE 2.2.5-2. SUMMARY OF UTILITY BOILER FGD PROCESSES [203]

	Percent of Capacity by Process Type		
Process	Operational	Construction	Total Market
Lime/Limestone	91	89	90
Sodium Carbonate	3	3	3
Magnesium Oxide	1	0	-
Double Alkali	2	5	4
Wellman-Lord	3	3	3

TABLE 2.2.5-3 SUMMARY OF INDUSTRIAL BOILER FGD PROCESSES [204]

	Percent of Capacity by Process Type		
Process	Operational	Construction	Total Market
Sodium Scrubbing	76	54	72
Double Alkali	11	35	16
Lime/Limestone	3	-	2
Others	10	11	10

TABLE 2.2.5-4. PERFORMANCE DATA FOR OPERATING SODIUM SCRUBBING SYSTEMS[205]

Installation/location	Sorbent	Fuel		Start-up Date	No. of FGD Units	SO_2 Inlet (ppm)	Percent Removal	Waste Disposal
		Type[1]	%S					
Alyeska Pipeline Valdez, Alaska	NaOH	O	<0.1	6/77	1	150	96	oxidation/dilution
American Thread Martin, NC	Caustic waste	C	1-1.5	1973	2	500	70	pond
Belridge Oil McKittrick, CA	NaOH	O	1.1	6/78	2	500	90	waste water treatment
Canton Textiles Canton, GA	Caustic waste	C	0.8	6/74	1	500	70	pond/waste treatment
Chevron Bakersfield, CA	Na_2CO_3	O	1.1	7/78	3	700	90	pond/waste treatment
FMC Green River, WY	Na_2CO_3	C	1	5/76	2	800	95	pond
General Motors Dayton, OH	NaOH	C	0.7-2.0	9/74	2	1.43#/10^6 BTU	86	clarify/adjust pH/ to sewer
General Motors Pontiac, MI	NaOH	C	0.8	4/76	2	—	—	combine with ash/ landfill
General Motors St. Louis, MO	NaOH	C	3.2	1972	2	2000	90	oxidize/neutralize/ discharge
General Motors Tonawanda, NY	NaOH	C	1.2	6/75	4	1#/10^6 BTU	90	combine with ash/ landfill
Georgia Pacific Orosett, AK	Caustic waste	B,C,O	1.5-2	7/75	1	500	80	to city sewers
Getty Oil Bakersfield, CA	Na_2CO_3	O	1.1	6/77-12/78	6	600	90-96	pond
Great Southern Cedar Springs, GA	Caustic waste	B,C,O	1-2	1975	2	1000	85-90	ash pond
ITT Rayonier Fernandina, FL	NaOH	B,O	2-2.5	1975	2	1200	80-85	to paper process
Kerr-McGee Trona, CA	Na_2CO_3	O	0.5-5	6/78	2	—	98	pond
Mead Paperboard Stevenson, AL	Na_2CO_3	O	1.5-3	1975	1	1500	95	to paper process
Mobil Oil San Ardo, CA	Na_2CO_3/NaOH	O	2-2.5	1974	28	1500	90	pond
Nekoosa Papers Ashdown, AK	NaOH	C	1-1.5	2/76	2	600	90	waste treatment
Northern Ohio Sugar Freemont, OH	NaOH	C	1	10/75	2	—	—	pond
St. Regis Paper Cantonment, FL	NaOH	B,O	<1	1973	1	—	80-90	clarification/ aeration
Texaco San Ardo, CA	NaOH	O	1.7	11/73	32	1000	73	pond/wells/softening and resuse
Texasgulf Granger, WY	Na_2CO_3	C	0.7	9/76	2	860	90	pond

(1) C=coal
O=oil
B=bark

C. Applicability--Sorbent costs and disposal of the waste liquor are the major limitations to this process. Because of the sorbent cost, Na_2CO_3 is \$60/ton f.o.b., and about \$90/ton delivered a distance of 1,000 miles,[206] applications of this process may become centered near large raw material sources which are in the Western part of the U.S. However, a significant number of industrial installations produce a sodium based waste stream (e.g. paper mills) that can be used as the sorbent such that the process may continue to be applied throughout the U.S.

Waste liquor disposal is the other major limiting factor with regard to the application of this technology. The majority of sodium scrubbing systems in use today are located in the California oil fields where the aqueous wastes are disposed of in evaporation ponds. If wastes from this system cannot be treated in existing waste water treating facilities or used as a process make-up stream, costs associated with achieving a zero discharge water system will more than likely limit the system's application due to economic reasons.

Sodium scrubbing, because it is simple both chemically and mechanically, can be applied to boilers of varying size and type. At present, sodium systems are being employed on industrial boilers ranging in size from about 4 to 100 MWe with satisfactory results. The process has been applied to both stoker-fed and pulverized coal boilers. The FGD system at the G.M.C. Delco Moraine plant in Dayton, Ohio controls SO_2 emissions from two stoker-fed, coal-fired boilers. The SO_2 removal efficiency is reported to be 85 percent and the average operability of the system is 95 percent.[206]

D. Factors affecting performance--

1) Design and operating considerations - One of the major factors to consider in system design is whether or not to remove fly ash prior to SO_2 removal. If the fly ash is removed then the absorber can be either a packed tower or a tray tower. If fly ash is not removed a venturi scrubber may be used for both particulate and SO_2 removal but this penalizes the process by requiring a higher absorber pressure drop. The fly ash contributes to solids buildup at the wet-dry interface and causes erosion of pipes, pumps,

spray nozzles and scrubber internals. The dramatic effect of pressure drop can be seen in Figure 2.2.5-2 representing Shawnee data for the prototype venturi scrubber.

Because the sodium alkali is very reactive the design L/G ratios can be low. In 1972 a series of tests were carried out at the Shawnee Test Facility using sodium carbonate. The effects of L/G are shown in Figure 2.2.5-3 for a marble bed scrubber, and for a TCA operating with no internals (*i.e.*, as a spray tower) in Figure 2.2.5-4. Efficiencies of 99 percent were realized in a TCA with a normal three-bed configuration.[210]

2) Fuel Variations - Variations in the fuel characteristics will affect the design of sodium scrubbing systems in the same manner they affect the design of the other wet FGD systems. The most important fuel characteristics with regard to FGD systems design are: sulfur content, chloride content, and alkalinity, and fuel heating value. Effects of these variables on process design are discussed in Section 2.2.1.1.

3) Ambient Variations - FGD systems are essentially independent of ambient variations. However, as with all wet systems, extreme cold can adversely affect the operation of an FGD process. Cold weather effects can, of course, be accounted for in the system design by providing heat traced lines or by enclosing the FGD system within a building.

4) Maintenance Problems - Problems encountered with sodium systems should be generally low. The system is simple to operate and does not require a complex sorbent regeneration process. Using a clear solution rather than a slurry minimizes potential plugging or scaling problems.

Some operating problems have, however, been reported in the EPA Industrial Boiler FGD Survey by system users. Most of the problems experienced to date concern corrosion of the scrubber and pH monitoring equipment, and erosion in systems that remove both SO_2 and flyash. A potential cure for these problems is the use of more exotic grades of materials as is commonly done in the utility industry.

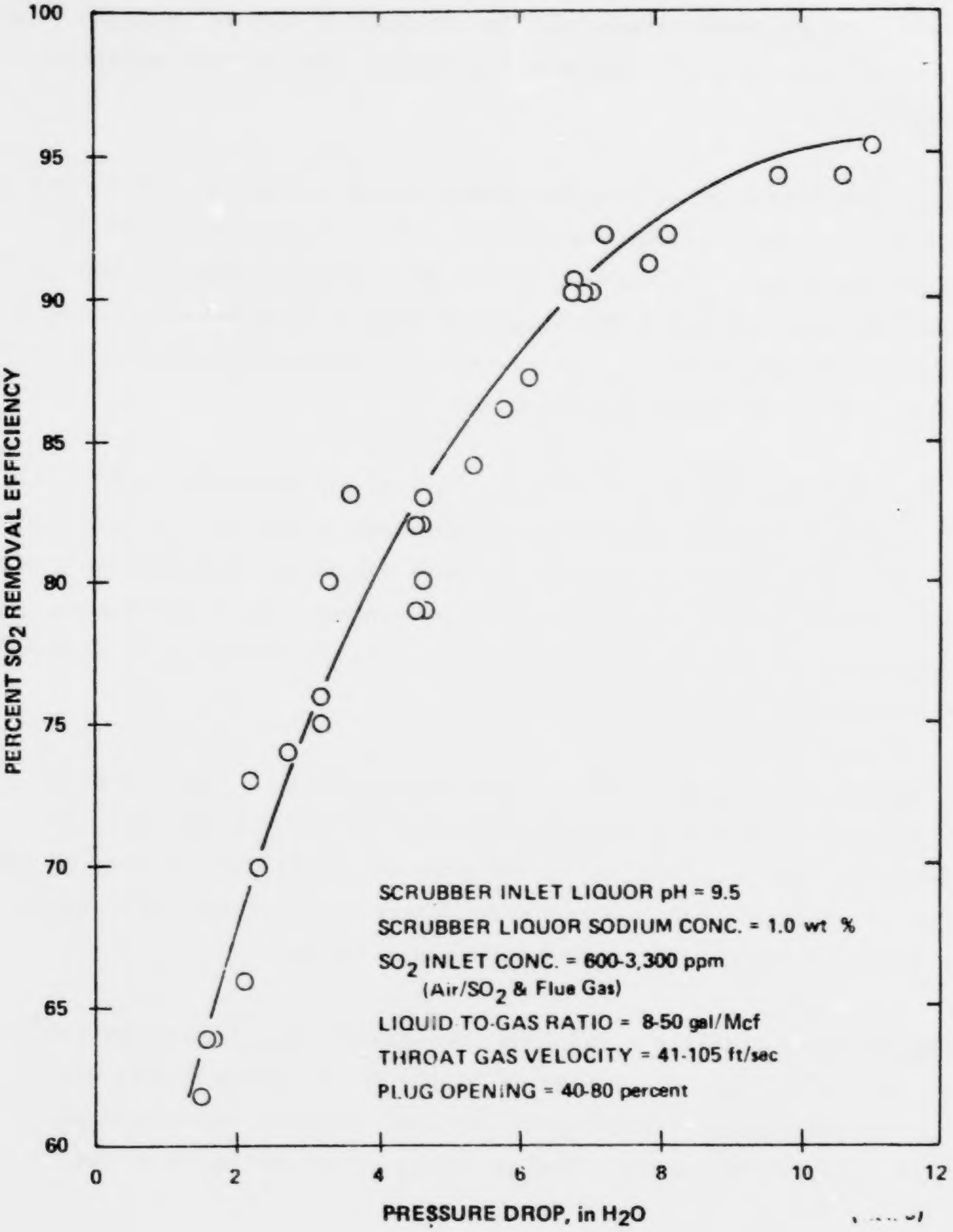


Figure 2.2.5-2. Effect of pressure drop on SO_2 removal efficiency - venturi with sodium carbonate (10 MW size).[207]

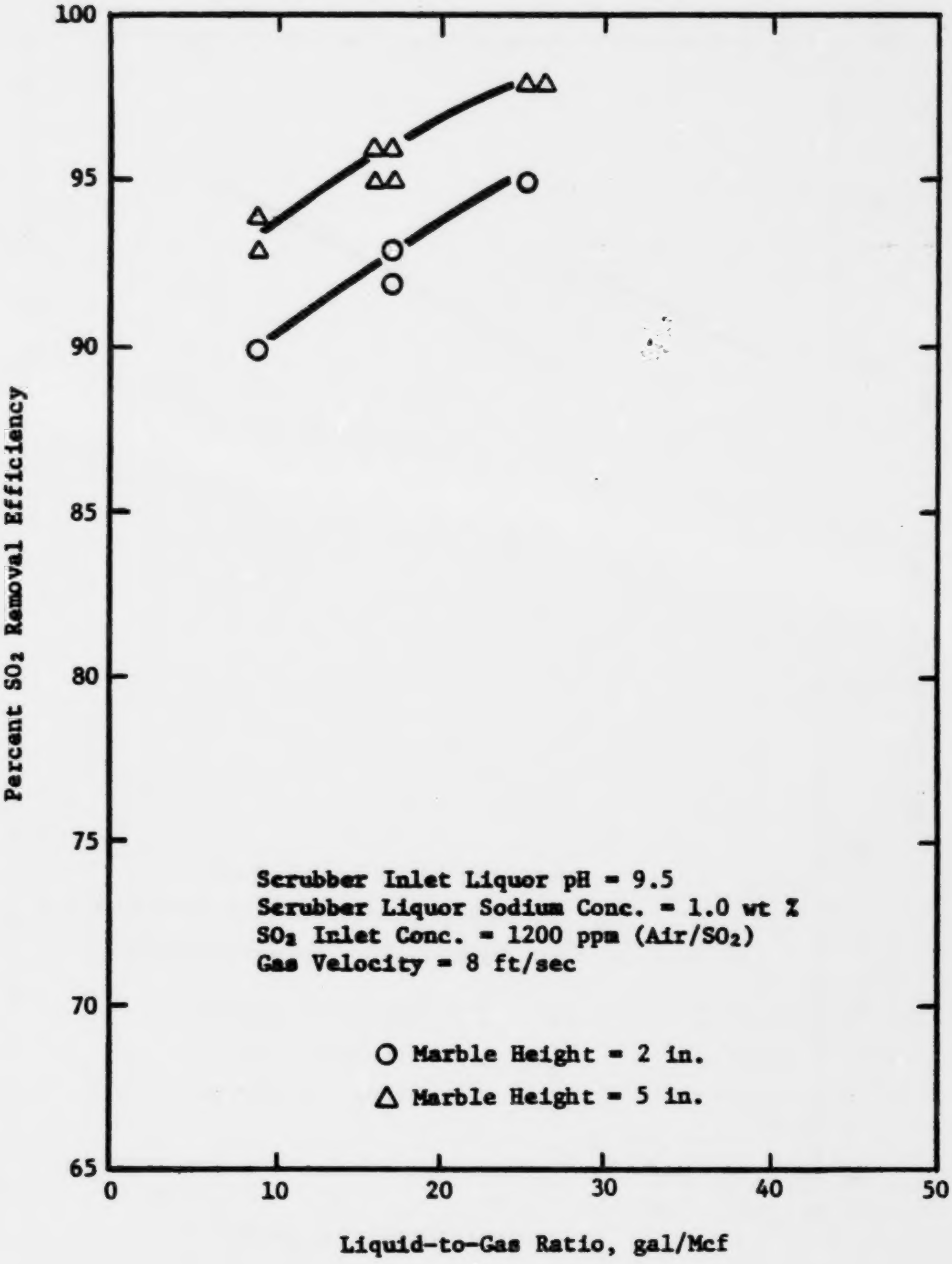


Figure 2.2.5-3. Effect of liquid-to-gas ratio on SO_2 removal efficiency.[208]

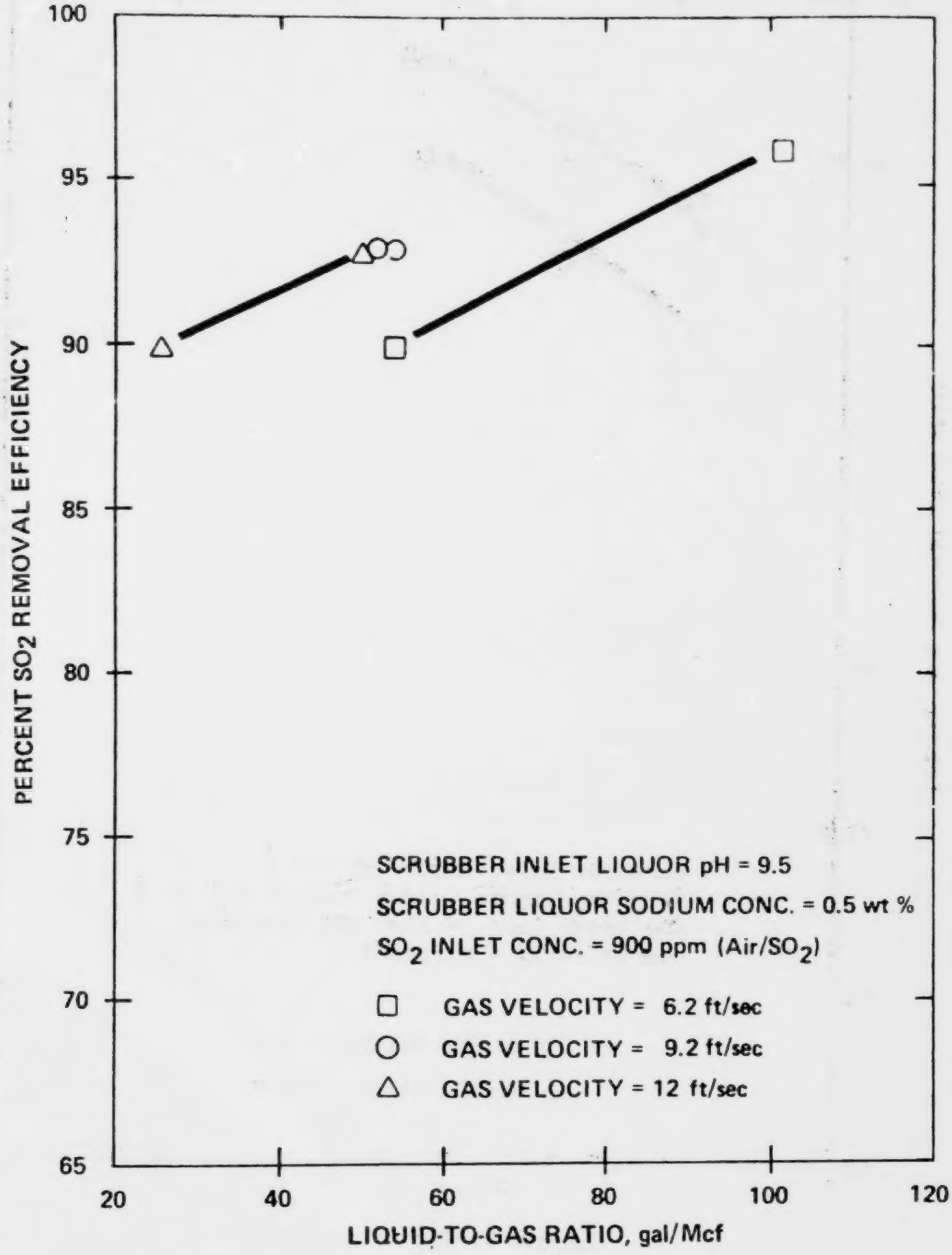


Figure 2.2.5-4. Effect of liquid-to-gas ratio on SO_2 removal efficiency - TCA (no spheres) with sodium carbonate (10 MW size).[209]

E. Retrofits--Sodium scrubbers should be well suited for retrofits. The system can be easily modularized and a relatively small amount of equipment is needed for the process. Effects on the boiler should be minimal. Retrofits may be limited by the availability of raw scrubbing materials and the cost of waste disposal.

2.2.5.2 System Performance--

A. Emission reduction--Although there are many sodium scrubbing systems in operation today, there is a lack of performance data for these systems with regard to their SO_2 removal capabilities. System operators, have, however, reported some data on process operations in the EPA Industrial Boiler FGD Survey. These SO_2 removal data, which are summarized in Table 2.2.5-4, indicate the SO_2 removal capabilities for the currently operating sodium scrubbing systems to be about 90 percent.

B. Availability/Reliability--Overall reliability of sodium scrubbing systems applied to industrial boilers has generally been quite high. Although no quantitative reliability/availability/operability indices are available from the EPA Industrial Boiler Survey prior to the 4th quarter of 1978, the majority of user responses indicated that the systems have been operating well with no problems being experienced, and hence little boiler downtime has been attributed to scrubber reliability problems.

An examination of the 4th quarter 1978 data shows that of the 22 industrial boiler installations that have operating sodium scrubbing systems, 15 reported quantitative reliability or operability indices that ranged from 89 to 100 percent with an average of 97.8. Of the 15 responses, 9 reported a 100 percent reliability/operability and all but two reported reliabilities of greater than 95 percent.[211]

Of the seven installations that did not report quantitative reliability indices, two reported that the FGD system had no problems, two reported erosion/corrosion problems, one was down for reconstruction, one was having mechanical problems with pump packings, and one system had no reported comments.

C. Impacts on boiler--Major effects of a sodium system on a boiler performance would be 1) boiler derating due to energy required to run pumps, fans, and if necessary, flue gas reheat, and 2) boiler load reduction due to scrubber downtime assuming no bypass of flue gas. The energy penalty for sodium scrubbing systems has been estimated to be about 2 percent of the net heat input to the boiler (see Section 5). Boiler load reduction due to scrubber downtime should not be significant due to the high reliability of sodium systems.

D. Additional maintenance requirements--Since sodium systems are a clear liquor process excessive maintenance problems are not expected in the absorber section. Manpower needs have been reported to vary from 0.25-1.0 man/shift for the smaller units (10-40 MW) to 3 men per unit for the larger systems (125 MW).[212,213]

2.2.6 Processes Under Development

At present, there are some 100 FGD systems in various stages of development. This includes systems in the very early stages of development to systems whose development efforts have ceased. A description and evaluation of all these systems is outside the scope of this project. However, some of these developing FGD systems may prove to be advantageous for near-term application to industrial boilers. Consequently, this section presents brief evaluations of the following FGD technologies that have not yet been operated in a commercial system but that have the potential for future commercial applications:

1) Dry Scrubbing
2) Citrate/Phosphate Buffered Absorption
3) Bergbau-Forsehung/Foster Wheeler

4) Atomics International Aqueous Carbonate
5) Shell/UOP
6) Chiyoda 121

2.2.6.1 Dry Scrubbing--

Several types of dry scrubbing processes are currently under development by EPA and FGD process vendors. Dry scrubbing systems that appear to be particularly applicable to industrial boilers include spray drying of a lime or sodium sorbent, and firing of a pellitized limestone/coal mixture in a stoker-fired boiler. This evaluation will focus on spray drying control systems since they are currently being installed on two commercial sized industrial boiler applications. The pellitized limestone/coal scrubbing process will only be briefly discussed due to its early development status.

In this process, the limestone/coal pellets are fired as ordinary fuel in stoker boilers. The SO_2 formed during combustion reacts with the limestone present in the fuel pellets to form calcium sulfite and calcium sulfate salts. The majority of calcium salts remain in the ash bed and are discharged from the boiler along with the bottom ash. This system does produce an increase in the boiler's particulate emissions which may affect the design of the fly ash control equipment.

Preliminary results from laboratory testing at EPA indicated that SO_2 removal efficiencies of 70-80 percent may be achievable with this type of control technology. These encouraging results were obtained on an Illinois No. 6 coal (about 3.5 percent sulfur) using fuel pellets with a calcium/sulfur ratio of about 7. Additional work has been conducted to optimize the structural properties of the pellets to obtain a fuel with the same handling characteristics as raw coal. This work has resulted in pellets with a calcium/sulfur ratio of about 3. Tests are currently being conducted to evaluate the combustion and SO_2 removal ability of the improved pellets.

Preliminary economic estimates have been prepared which indicate this technology will cost about $15/ton of coal. This cost compares very well with the annual costs estimated for other industrial boiler FGD systems which range from about $23 to $27 per ton of coal burned assuming a 200×10^6 Btu/hr boiler. This control technology, when developed, has the potential for providing a low cost SO_2 removal option that may be rather easily applied to both new and existing industrial boilers.

A. System description-- In a spray drying process, a slurry of soda ash or lime is used to remove SO_2 from boiler flue gas. The spray dryer product is a dry mixture of sodium or calcium salts and unreacted sorbent which can be collected with flue gas fly ash for disposal.

Flue gas enters the spray dryer at air preheater exit temperatures, generally between 250 and 300°F. In the spray dryer the gas is contacted by high speed centrifugal atomizers and driven outward in cross flow to the flue gas. Flue gas velocity in the dryer vessel is on the order of 5 ft/sec. The dryer vessel is sized for approximately five to ten seconds residence time.

In the scrubber, gaseous SO_2 is sorbed into the liquor where it reacts with sorbent material to form sulfite salts, as indicated in the following Reactions:

$$SO_2\ (g) + Na_2CO_3(aq) \rightarrow Na_2SO_3(aq) + CO_2(g) \qquad (2.2.6\text{-}1)$$

$$SO_2\ (g) + CaO(s) + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O \qquad (2.2.6\text{-}2)$$

In addition to these primary reactions, sulfate salts will be produced by the following reactions:

$$Na_2SO_3(aq{+}s) + \tfrac{1}{2}O_2(g) \rightarrow Na_2SO_4(aq{+}s) \qquad (2.2.6\text{-}3)$$

$$SO_3(g) + Na_2CO_3(s) \rightarrow Na_2SO_4(s) + CO_2(g) \qquad (2.2.6\text{-}4)$$

$$SO_2(g) + CaO(s) + \tfrac{1}{2}O_2(g) + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s) \qquad (2.2.6\text{-}5)$$

A typical product mixture formed by these reactions when using a sodium sorbent is approximately 60 weight percent sodium sulfite, 20 percent sulfate, and 20 percent excess carbonate.[214] Expected composition of the waste material from a lime based spray drying system is about 55 percent calcium sulfate, 30 percent sulfite, and 15 percent $CaCO_3$ and lime inerts. This estimate is for a system that recycles a portion of the collected sorbent fly ash mixture in order to improve the overall lime utilization.[215]

Liquid to gas (L/G) ratios for spray drying are typically 0.2 to 0.3 gal/1,000 scf. This low liquid rate is insufficient to saturate the gas. Gas exit temperatures are typically in the 150 to 200°F range to provide a safe margin against water condensation. Spent reactant is entrained in the flue gas as dry particulate material.[216]

SO_2-clean flue gas exits the spray dryer and is routed to a normal particulate collection device such as an ESP or baghouse, where spent reactant and fly ash are removed for disposal. Systems using a baghouse for particulate removal report additional SO_2 sorption occurring in the baghouse as will be discussed later. Care must be taken to maintain flue gas temperature well above saturation[217] at this point to avoid condensation on the solids collection device surfaces.

Accessory equipment consists of liquor preparation and dry waste disposal facilities. In general, liquor preparation facilities include dry storage, a liquor mix tank, and associated liquor tankage and pumps. Facilities for handling the collected spray dryer product and fly ash and transporting them to the ultimate disposal site are similar to those normally associated with baghouse or ESP collection devices. A generalized process flow diagram for a typical spray drying scheme is shown in Figure 2.2.6-1.

B. Development status--Spray drying technology for removing SO_2 from boiler flue gas is still under development, although spray dryers have been used in various industrial applications for many years. Development and

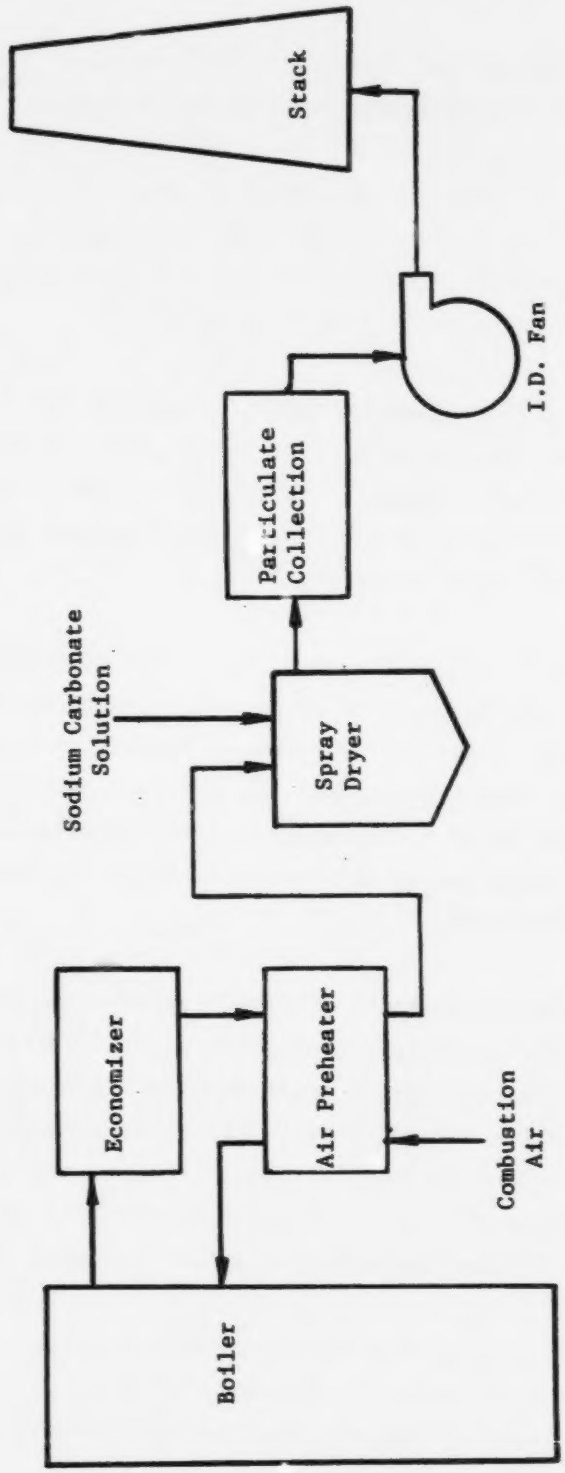


Figure 2.2.6-1. Simplified Flow Diagram for Spray Drying Process.

pilot plant installations have been conducted in the United States by Atomics International/Wheelabrator Frye, Joy Manufacturing Company/Niro Atomizer, Babcock and Wilcox, and Carborundum. Additional process vendors which have announced dry SO_2 control system development efforts include Buell/Envirotech, Ecolaire, Micropul, and Research Cottrell. Other process vendors may also be pursuing dry scrubbing systems due to the widespread interests on the part of potential utility and industrial users. Major U.S. development efforts have included pilot-scale testing at Southern California Edison's Mohave Station, Basin Electric's Leland Olds Station, and Ottertail Power's Hoot Lake Station. In addition, it has been reported that Koyo Iron Works is undergoing development of a sodium based spray dryer system in Japan.[218]

To date, although there are no spray drying SO_2 control systems currently operating on either an industrial or utility application, five contracts have been awarded for construction of commercial sized systems ranging in size from 22,000 SCFM (about 11 MWe) up to about 575 MW. Two of these systems are industrial boiler applications. Table 2.2.6-1 summarizes the status of these five systems. These activities indicate a significant amount of interest in this rapidly developing control technology.

TABLE 2.2.6-1 SUMMARY OF COMMERCIAL DRY SCRUBBING APPLICATIONS[219,220,221]

System/Location	Vendor	Sorbent	Size	Start-up Date
Strathmore Paper Co. Woronoco. MA	Mikropul	Lime	22,000 SCFM (11 MWe)	5/79
Celanese Cumberland, MD	Wheelabrator-Frye/ Rockwell Int.	Lime	50,000 SCFM (25 MWe)	12/79
Basin Electric Co. Antelope Valley #1	Joy/NIRO	Lime	440 MWe	4/82
Basic Electric Co. Laramie #3	Babcock & Wilcox	Lime	575 MWe	4/82
Otter Tail Power Coyote Unit #1	Wheelabrator-Frye/ Rockwell Int.	Na_2CO_3	410 MWe	Spring'81

C. Design considerations--Based upon pilot unit test results, 90 percent SO_2 removal can be achieved using either lime or sodium based sorbents. Stoichiometric ratios of 2.3-3.0 were required for lime operations whereas stoichiometric ratios of only 1.0-1.2 were required to achieve the same SO_2 removal for sodium operations. It has also been reported that 90 percent SO_2 removal may be achieved with a stoichiometric lime requirement of 1.3-1.7 by recycling some of the unreacted sorbent.[222] These reported values are for relatively low sulfur coal operations. It still remains to be demonstrated whether or not a lime based spray dryer system will be able to achieve high SO_2 removal efficiencies when applied to a high sulfur coal. A sodium based system should be able to achieve high SO_2 removals on high sulfur coals due to its higher reactivity.

Process control of the spray dryer feed solution is basically simple and straightforward. However, the amount of scrubbing solution added to the spray dryer must be carefully controlled as too much will result in condensation in downstream equipment and not enough will prevent attaining the required SO_2 removal efficiency.

The primary reaction of SO_2 with sorbent material in the spray dryer appears to occur in the aqueous phase. SO_2 removal and sorbent utilization are, therefore, enhanced by increasing the exposure time of liquid droplets to flue gas SO_2. This liquid phase residence time in the spray dryer can be lengthened by increasing the ratio of liquid to flue gas (L/G) entering the scrubber. An increase in L/G increases aqueous phase residence time not only because the mass quantity of water to be evaporated is larger, but because adiabatic evaporation of the increased moisture content lowers the flue gas outlet temperature. As the flue gas temperature decreases toward its adiabatic saturation temperature, the driving force for evaporation also decreases. Higher L/G ratios also result in better gas-liquid contact because higher liquid feed ratios result in more spray droplets.

For a given flue gas inlet moisture content and temperature, an upper limit for L/G ratios is set by the temperature requirement for the outlet flue gas. Flue gas from the spray dryer must be approximately 30-50°F above its adiabatic saturation temperature in order to avoid condensation on particulate removal device surfaces and to avoid plume opacity due to condensation.

For a given inlet moisture content, a higher flue gas inlet temperature will result in the capability of evaporating more water. For example, a flue gas entering a spray dryer at 250°F with a humidity of 0.06 lb water per lb dry gas would have an adiabatic saturation temperature of approximately 124°F. Adiabatic evaporation to within 50°F of saturation would result in an outlet humidity of approximately 0.08 lb H_2O/lb dry gas. This humidity increase across the spray dryer corresponds to an L/G ratio of 0.17 gal/1000 scf. Increasing the inlet temperature to 290°F at the same 0.06 lb H_2O/lb dry gas inlet humidity raises the adiabatic saturation temperature to approximately 127°F, and raises the humidity at a 50°F approach to saturation to approximately .09 lb H_2O/lb dry gas. This raises the maximum L/G ratio possible to 0.26 gal/1000 scf.

For this reason, industrial boilers appear to be particularly well suited for spray drying applications since they typically do not have as sophisticated of a flue gas heat recovery system as utility boilers and tend to have hotter flue gas exit temperatures. For the cases considered, the industrial boiler flue gas temperatures ranged 350 to 400°F instead of 250-300°F which is typically found on utility boilers.

Since spray dryer FGD systems have significant amounts of unreacted sorbents in the exit gas, the potential exists for further reaction with flue gas SO_2 in downstream equipment. Since the temperature and gas/solid relative velocities are low in the outlet from a spray dryer, no significant reaction of SO_2 with unreacted sorbent can be expected in the subsequent duct work. However, collection of the solid material in a baghouse offers a high gas/solid relative velocity as well as a high solid residence which are both conducive to SO_2-sorbent reactions.

Data reported from various test installations appears to vary somewhat with regard to the amount of SO_2 absorption occurring in a baghouse particulate collection device. Results of one pilot unit indicated that up to 20 percent of the system's total SO_2 removal occurred in the baghouse when sodium was the sorbent and up to 10 percent occurred when lime was the sorbent. Results from another pilot system which was investigating the performance of lime as the sorbent indicated that virtually no SO_2 absorption was observed in the baghouse.[223]

Design considerations directly related to the performance of a spray dryer-based FGD system are the operation and maintenance of the system. The successful operation of a spray dryer FGD system is greatly affected by the process control instrumentation. Unlike wet scrubbing systems in which SO_2 removal can be controlled by some gross recycle liquor physical parameter such as pH, this FGD system must be controlled by a feedback loop from continuous monitoring of outlet SO_2 concentration and temperature. Increases in spray dryer outlet temperatures from set point would result in increased liquor flow rates to the scrubber and vice versa. Increases or decreases in system outlet SO_2 concentrations would result in a corresponding increase or decrease in sorbent feed to the scrubbing liquor mix tank. In such a control scheme, an undetected SO_2 monitor drift or error could result in failure to meet control requirements or waste of sorbent, depending upon the direction of error. Error in outlet temperature measurement could result in excessive L/G ratios and consequent condensation on downstream particulate control equipment.

During one of the pilot plant installations, a spray dryer was deliberately operated under upset conditions to examine the effect of condensation on baghouse performance. After the system had returned to normal operating conditions for one hour it was inspected and no changes were found on the baghouse and bags when compared to previous inspections. This was because the exit gases at normal temperatures dried the wet deposits to the point that they were removed in the normal bag cleaning cycle. This upset condition was for a short duration (15 minutes) so it is still unclear what the effects of a long term upset will be.[224]

A spray dryer-based FGD system offers maintenance and operation advantages over other systems in 1) the high turndown ratios which can be achieved and 2) the anticipated low maintenance requirements and high reliability of the major process equipment. Turndown of the spray dryer system should be relatively simple and high turndown ratios should be possible. As gas rate decreases, liquid feed rate to the scrubber is decreased proportionally. Turndown ratios of up to 4 to 1 have been demonstrated in a single scrubber with a single atomizer with no decrease in SO_2 removal efficiency. The benefit of increased gas residence time appears to maintain SO_2 removal efficiencies at lower flow rates. In larger installations where multiple spray dryer vessels would be employed, dampers could be employed to remove whole vessels from service to increase the possible turndown ratio. The major limit to system turndown is the onset of condensation in downstream equipment due to greater heat losses at the significantly lower flue gas flow rate.[225]

The spray dryer scrubber could offer some advantage over wet scrubbers in maintenance and reliability of process equipment. Because of the much lower L/G ratio and the relatively small head requirements, pump maintenance requirements for this system should be significantly lower. Scrubber vessel problems such as scaling or erosion should be avoided, since wet particles should never come into contact with vessel walls. While some deposition of dry material on scrubber walls has been observed in demonstration tests, occasional mechanical rapping has been found to adequately control buildup.[226]

Since the energy for atomization comes from a spinning disc rather than from a nozzle constriction, atomizer erosion problems should be minimal. However, due to their high speed of operation (up to 18,000 rpm) it can be expected that the atomizers will require frequent periodic maintenance. Due to the simplicity of removal of a single atomizer and the fact that multiple atomizer arrangements can be employed in spray dryers in a large installation, failure or removal from service of one atomizer should have a minimal effect on overall system performance. In October 1975 tests on a 7-foot test scrubber by Bowen Engineering, the equipment supplier for the Atomics International process, indicated that removal efficiency in a three atomizer spray dryer could be maintained when removing flow from one atomizer by adding this flow to the other two atomizers.[227]

D. Applicability to industrial boilers--There are no apparent technical constraints in applying spray drying technology to small industrial boilers. In fact, due to the relative simplicity of the system and its claimed high reliability, this technology may prove to be a very desirable option for small industrial applications.

Possible concerns with applying this technology to industrial boilers are in the following areas: 1) the relatively large diameters of the spray drying vessels, 2) the possible detrimental effects on downstream particulate collection equipment from system upsets, and 3) lack of data concerning the system's SO_2 removal ability for high sulfur eastern and midwestern applications. Another concern, which is not unique to this system, is the problem of solid waste disposal.

Condensation effects on downstream particulate control equipment and system performance on various coals under actual operating conditions are questions that will be resolved after additional operating experience with the system is gained. Industrial boiler spray drying systems at Strathmore Paper Co. and Celanese which are expected to be operational in late 1979 will be the first commercial applications of the process.

E. Summary--The major advantages claimed for using a spray drying system include: 1) less equipment requirements than a dry system, 2) production of a dry rather than wet waste material, 3) reported high reliability of system components, 4) lower system costs, and 5) reduced energy requirements

The major unresolved issue appears to be the effect of auxiliary equipment such as control instrumentation on overall system reliability. Although the major process equipment has proven to be quite reliable in other applications, it remains to be seen how precisely the spray drying system can be controlled to avoid condensation in subsequent particulate collection equipment under low temperature, high moisture boiler flue gas conditions.

2.2.6.2. Citrate Buffered Absorption--

The citrate flue gas desulfurization process is a regenerable process which removes SO_2 from flue gases by absorption in an aqueous sodium citrate solution. The absorbed SO_2 is converted to elemental sulfur in subsequent processing steps. The U.S. Bureau of Mines pioneered the development of the Citrate Process with developmental work also being done by Arthur G. McKee and Company, Peabody Engineering, and Pfizer, Inc.

A. System description[228]--Flue gas is quenched and prescrubbed with water to remove HCl, SO_3, and residual fly ash. SO_2 is removed by a buffered aqueous solution in an impingement tray or packed bed absorber. The flue gas may be reheated to 175°F before discharge to the atmosphere. The primary reactions occurring in the absorber are as follows:

$$SO_2 + H_2O \rightarrow HSO_3^- + H^+ \qquad (2.2.6\text{-}6)$$

$$HCit^= + H^+ \rightarrow H_2Cit^- \qquad (2.2.6\text{-}7)$$

The citrate acts as a buffer maintaining a pH of 3-5 throughout the system. The regeneration reaction requires a pH in this range.

Absorber liquor is reacted with H_2S in several reactors placed in series to form elemental sulfur by the following overall reaction:

$$SO_{2\,(\ell)} + 2H_2S \rightarrow 3S + 2H_2O \qquad (2.2.6\text{-}8)$$

The sulfur is concentrated and separated from the regenerated liquor. H_2S for the regeneration reactors can be produced by reacting reducing gas with two-thirds of the product sulfur from the process.

Some sulfite oxidation occurs and results in the formation of sodium sulfate which must be removed from the system by crystallization as Glaubers salt ($Na_2SO_4 \cdot 10H_2O$). The amount of sulfite oxidation in the systems has been reported to vary from one to four percent. The low oxidation rates are attributable to the presence of thiosulfate ions which are formed in the regeneration step.

Figure 2.2.6-2 presents a simplified flow diagram of this process which highlights the interaction of the various processing areas. This process is made up of many components and is consequently relatively complex.

Process equipment used in addition to the prescrubber and absorber are: regeneration reactors, sulfur flotation and melting equipment, sulfate purge crystallization equipment, and H_2S generation equipment. Except for the regeneration reactors and H_2S generation equipment, the process operations have been well demonstrated in other applications. Integration of all equipment and demonstration of system reliability and ability to follow boiler operations will be demonstrated on an industrial boiler (64 MW) located at St. Joe Minerals.

B. Development status[229]--The Citrate Process was pioneered by the Bureau of Mines with developmental work also done by Arthur G. McKee, Peabody Engineering, and Pfizer, Inc. The Bureau of Mines operated a 1,000 scfm pilot unit on a lead sintering furnace tail gas for about two years. Pfizer, Peabody, McKee treated 2,000 scfm from a coal-fired industrial boiler for about seven months in 1974. A Bureau of Mines/EPA co-funded demonstration unit on 120,000 scfm of flue gas from a coal-fired industrial boiler is scheduled to start up in mid 1979. This unit is at St. Joe Minerals zinc smelter in Pennsylvania.

The primary technical hurdles that remain for the Citrate Process are H_2S generation from a hydrogen source other than methane, integration of an H_2S generation unit with the process, and a more complete understanding of the chemistry of the regeneration reaction. The process has been totally integrated except for producing H_2S from a coal based reducing gas. H_2S was produced from methane in the pilot plant installations.

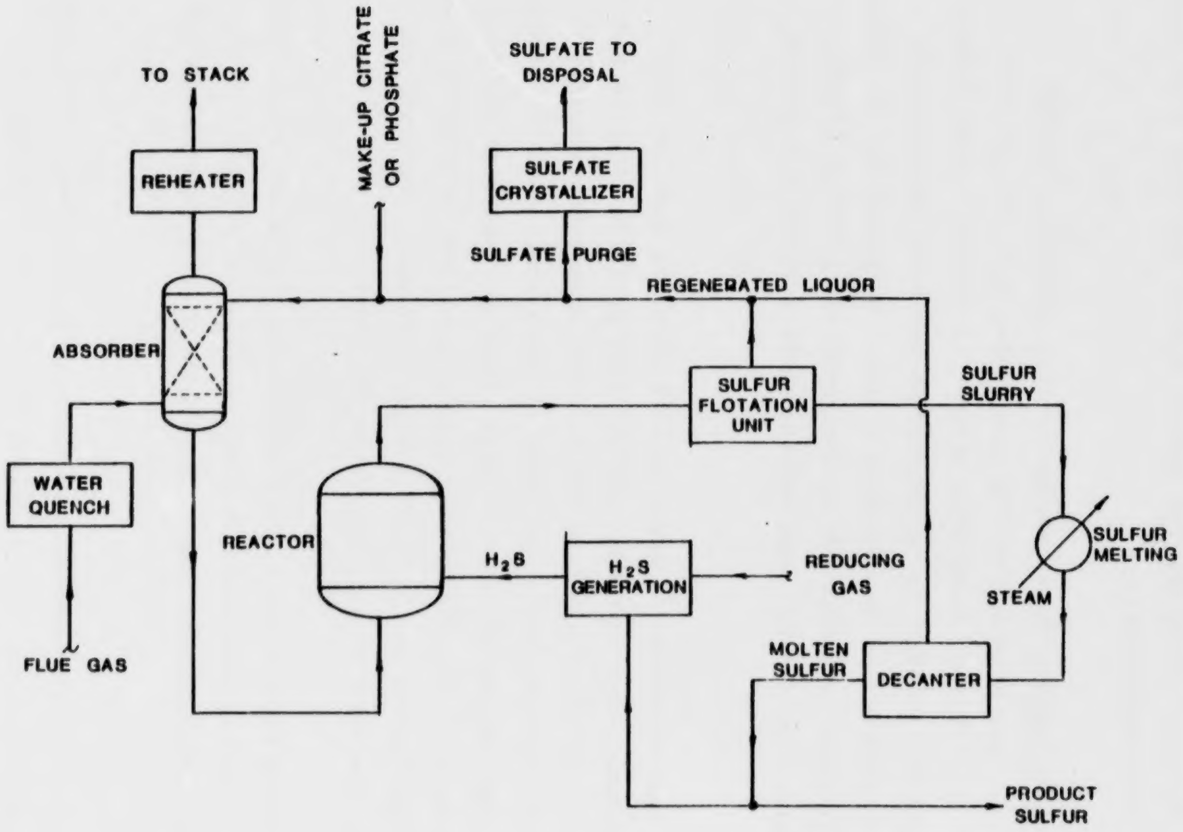


Figure 2.2.6-2. Simplified Flow Diagram for Citrate/Phosphate Process.

Currently, no data can be obtained for this process except for the earlier pilot plant test runs. However, the EPA demonstration of this system at St. Joe Minerals should provide data to answer questions concerning process chemistry and operability.

C. Design considerations--Major design factors affecting the SO_2 absorption portion of this process are flue gas temperature, liquor circulation rate, and pH. Process developers attempt to optimize these factors to achieve a desired SO_2 removal (up to 99 percent has been reported from pilot units) while producing as concentrated a liquor as possible. An effluent liquor highly concentrated with SO_2 is desirable since a high liquor loading produces less solution for regeneration which results in a smaller regeneration unit.

Regeneration reactor design is based on both reaction rate and mass transfer considerations. One developer reports that reaction kinetics are limiting; the mass transfer of H_2S into the solution occurs more rapidly than reactions involving intermediates formed in the system. Another vendor reports that the regeneration reaction is mass transfer-limited in the first reactor where the ratio of H_2S concentration to aqueous SO_2 concentration is lowest, and kinetics-limited in the other two reactors where the ratio of H_2S concentration to aqueous SO_2 concentration is greater.[230]

The concentration of hydrogen sulfide in the gas fed to the regeneration reactors is another design consideration since it directly affects reactor size. Vendor specifications vary from a 39 percent to a 96 percent nominal H_2S gas stream.

Small bubbles of H_2S are desirable in the regeneration reactors to provide more liquid-gas interface. Pilot plants have used an open pipe or sparger generating large bubbles with a high tip-speed agitator shearing the large bubbles into tiny bubbles. The regeneration section will precipitate sulfur and, therefore, must be designed to minimize plugging problems.

An H_2S stream produced on site from reducing gas or hydrogen and sulfur will contain some elemental sulfur vapors. Potential condensation and solidification of these vapors upon contact with the liquor must be considered in determining the method of introducing H_2S gas to the regeneration reactors.[231]

Potential plugging at the entrance to the regeneration reactors could be reduced by several methods. Two parallel heat exchangers operated in a cyclic mode could be used. One would cool the H_2S gas and condense the sulfur vapors while steam was injected in the other to vaporize condensed sulfur. Alternately, a water quench could remove the sulfur and saturate the H_2S gas streams producing a sour water containing hydrogen sulfide.

Flotation with either kerosene or air has been examined for use in separating the precipitated sulfur from the regenerated liquor. The flotation unit should produce an essentially sulfur-free citrate/phosphate solution for recycle, and a sulfur stream containing as little liquor as practicable. Kerosene flotation yields a powdery damp sulfur of about 50 percent solids, while air flotation yields a pumpable slurry of about 7-15 percent solids. The sulfur slurry produced with air flotation contains more liquor than that produced by kerosene flotation. As a result, systems using air flotation will have more solution exposed to the sulfur melting and decanting steps. A potential for buffer and thiosulfate degradation exists at the operating temperature of about 280°F in the melting and decanting steps.[232] Kerosene flotation may affect product sulfur quality and may result in significant kerosene losses. An open vessel was used at the pilot plant with kerosene losses of about 90 lb/net ton of sulfur produced. A vapor recovery system on an enclosed vessel might reduce these losses to a more acceptable level.[233]

The sulfur melter is simply a steam heat exchanger that raises the temperature of the sulfur slurry out of the flotation cell to about 280°F. The equipment transporting the sulfur/liquor mixture to the melter and the melter itself must be designed for the appropriate slurry service and to minimize any potential for plugging. The decanter is a pressurized gravity flow separation unit. Molten sulfur is continuously or intermittently taken off the bottom while clear liquor is recycled to the absorber. All of the sulfur transfer lines must, of course, be heat traced to prevent solidification.

The sulfate purge section should present no unique design problems. A crystallizer, refrigeration unit, and Glauber's Salt ($Na_2SO_4 \cdot 10H_2O$) separator are used to remove the sulfates from the citrate/phosphate solution. The byproduct is then dried to produce anhydrous Na_2SO_4. The unit may be operated intermittently to control the sulfate concentration in the liquor. It should be noted that commercial sulfate removal units have not been operated at any of the pilot plants using the Citrate/Phosphate Process.

Hydrogen sulfide (H_2S) generation design considerations in the buffered absorption processes are different from most FGD processes that require H_2S because H_2S is produced externally by reacting elemental sulfur with hydrogen or reducing gas (H_2/CO). A gas stream with a high H_2S concentration is desirable for use in the regeneration section. A higher H_2S partial pressure results in a higher mass transfer rate and reduced reactor size. The H_2S concentration will, however, be determined mainly by the composition of the reducing gas or hydrogen fed to the H_2S generation unit rather than the process selected for H_2S generation.[234]

Plugging of the system by condensed sulfur has been one of the major operational problems with the H_2S generation facility. This unit must be designed so that condensation of sulfur vapors occurs only in knockout vessels included in the design.

The ability to turn down the unit or store H_2S is important for a workable process design. The H_2S generation unit must be able to follow the load swings of the boiler since hydrogen sulfide requirements are directly proportional to the quantity of SO_2 absorbed. Turn down capability has not been fully developed or demonstrated at any of the pilot plants.[235] The small amount of residual H_2S leaving the regeneration reactors will be collected and combusted in the boiler or an incinerator.[236]

D. Applicability to industrial boilers--Overall system complexity would appear to limit the applicability of this system to small industrial boilers. Since all of the liquor effluent from the absorber is routed to the regeneration section, this process is not well suited for a decoupled centralized regeneration.

The major concern with applying this system to an industrial boiler is with the sulfur reduction/H_2S generation portion of the process. Unless a stream of H_2 or CH_4 were available, the reductant would have to be produced from coal gasification which is still relatively undeveloped and would add to the overall process complexity. This process does, however, become rather attractive for applications to facilities with existing waste H_2S streams such as oil refineries. For these applications the H_2S stream could be introduced directly into the regeneration reactors, which would eliminate process problems associated with integrating an H_2S generator into the FGD operations.[237]

E. Summary--The unique features of the process are the reaction of H_2S gas with absorbed SO_2 in the regeneration step, the separation of precipitated sulfur by flotation, and the external generation of H_2S from elemental sulfur and reducing gas. Advantages of this process are production of sulfur as a by-product, and simple absorber operation which should result in good SO_2 removal. The major disadvantage of this process for application to industrial boilers is its overall complexity especially in regard to the H_2S generation section. The demonstration of this process at St. Joe Minerals will provide data to answer questions on process control, operability, and reliability.

2.2.6.3 Bergbau-Forschung/Foster Wheeler Process--

Flue gas SO_2 is adsorbed in a moving bed of char to form H_2SO_4 in the Bergbau-Forschung/Foster Wheeler (BF/FW) system. Saturated char is then heated to yield gaseous SO_2 and regenerated char. The SO_2 stream is reduced in the presence of anthracite coal to elemental sulfur. The adsorption and regeneration steps were developed by Bergbau-Forschung in Germany while the reduction step is proprietary technology of Foster Wheeler.

A. System description--The Bergbau-Forschung/Foster Wheeler process uses the concept of dry carbon adsorption of SO_2 and thermal regeneration of the carbon. A flow diagram for the process is presented in Figure 2.2.6-3. Flue gas from the boiler passes through dust collection equipment to remove the bulk of the particulate matter. The gas enters the adsorber at temperatures between 250-300°F. The flue gas passes through one or two beds of activated char where SO_2, SO_3, oxygen, and water vapor are adsorbed on the char. HCℓ is reported to pass through the adsorber without reacting the char. The following reactions take place in the presence of activated char:

$$SO_{2(g)} + \tfrac{1}{2}O_{2(g)} \xrightarrow{char} SO_{3(g)} \qquad (2.2.6\text{-}9)$$

$$SO_{3(g)} + H_2O_{(g)} \xrightarrow{char} H_2SO_{4(\ell)} \qquad (2.2.6\text{-}10)$$

The sulfuric acid formed remains in the pores of the char. The spent char is conveyed to a regeneration vessel where the char is mixed with hot sand at a mixture temperature of about 1200°F. The following reactions take place in the regeneration step.

$$H_2SO_{4(\ell)} \rightarrow H_2SO_{4(g)} \qquad (2.2.6\text{-}11)$$

$$H_2SO_{4(g)} \xrightarrow{1200°F} H_2O_{(g)} + SO_{3(g)} \qquad (2.2.6\text{-}12)$$

$$2SO_3 + C_{(s)} \xrightarrow{1200°F} CO_{2(g)} + 2SO_{2(g)} \qquad (2.2.6\text{-}13)$$

The carbon in Reaction 2.2.6-13 is supplied by consuming part of the char adsorbent. The char is recycled to the adsorber and the sand is recycled to a sand heater where it is reheated to about 1500°F.[238,239]

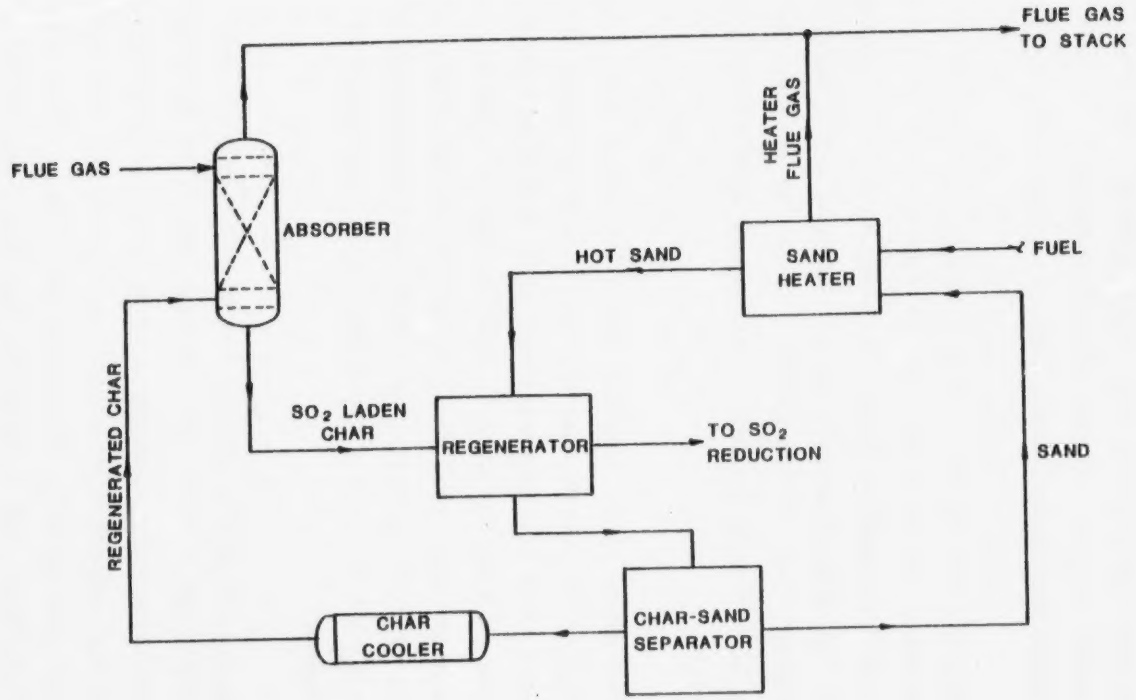


Figure 2.2.6-3. Simplified Flow Diagram for Bergbau-Forschung Process.

The off gas from the regenerator contains about 20 volume percent SO_2 and can be used to make sulfuric acid or sulfur. If the SO_2 was to be used to make sulfuric acid the gas would first be cooled and cleaned before being sent to the acid plant. If sulfur was to be produced the gas could be sent to the proprietary Foster Wheeler RESOX process. The gas would then be taken directly from the regenerator and reacted with crushed anthracite coal at 1100-1500°F to form elemental sulfur. Foster Wheeler generally includes the RESOX process in its Bergbau-Forschung/Foster Wheeler system designs.[240]

This process is characterized by its relative mechanical complexity as compared to wet FGD processes. It is a dry process that adsorbs SO_2 on solid carbon particles which must be transported to the regeneration processing area and then back to the adsorption area. Current designs use a series of conveyors and bucket elevators for this transfer. Since the regenerated char will be at approximately 1200°F this mechanical equipment must be designed for high temperature service.

Major equipment items are: 1) an adsorber which handles a moving bed of char, 2) a thermal regenerator which uses hot inert sand to raise the char temperature to about 1200°F for regeneration, 3) a char-sand separator, 4) a fluid bed sand heater to heat the sand to 1500°F, and 5) a RESOX sulfur production system which uses coal to directly reduce the regeneration off gases to sulfur. All of this equipment is solids handling equipment which in general, presents more operating and maintenance problems than gas-liquid handling equipment.

B. Development status--The BF/FW Process has been thoroughly and successfully tested at the bench, pilot, and prototype level in West Germany. In 1963, research and development began using a 20 scfm bench-scale system that included adsorption and regeneration steps. A process for the manufacture of char was developed concurrently in a facility which could produce approximately 5000 tons per year. As a result of the bench-scale work a pilot plant operating on a 1,750 scfm slip stream of flue gas from a 35 MW pulverized coal-fired boiler was operated from 1968-1970. A prototype 45 MW unit was installed by Bergbau Forschung in 1974 at the Kellerman Power Plant in Lunen, West Germany. The completely integrated process operates on an 88,000 scfm slip stream from a 350 MW coal-fired peaking service boiler. Off gas from the regenerator is treated in a modified Claus unit (BAMAG) to produce elemental sulfur. Testing at the prototype unit, which is sponsored by the West German government, is scheduled to end in late 1978.

The first fully integrated application of the BF/FW Process using RESOX was demonstrated by FW on a 47.50 MW coal-fired boiler at Gulf Power's Scholz Stream Plant. The unit consisted of a 20 MW adsorption section and a 27.5 MW reduction section (RESOX). The facility operated intermittently from August 1975 to May 1976. The unit experienced only limited success due primarily to mechanical problems or failures caused by Foster Wheeler in efforts to reduce costs. The Bergbau Forschung system at Lunen, West Germany, is the most successful system to date and is more representative of the process than the system at Scholz.[241]

C. Design considerations--The adsorber in the BF/FW Process is a unique gas-solid contacting device. The adsorber must be designed for gas flow as well as char flow. It is very important that both materials flow with even distributions and velocities throughout the adsorber. In the Bergbau design, a set of louvers perform several functions that influence the distribution and velocity of the gas and the char as shown in Figure 2.2.6-4. As a result, the most important aspect of the adsorber is the louver design.

The primary design criteria for the louvers are the angle of inclination at the gas inlet and outlet, the vertical spacing between the louvers, and the width of the louver measured in the direction of gas flow. These criteria have a direct bearing on:

- distribution of the char weight on the structural support columns,
- the uniform downward flow of char to insure plug flow in the beds,

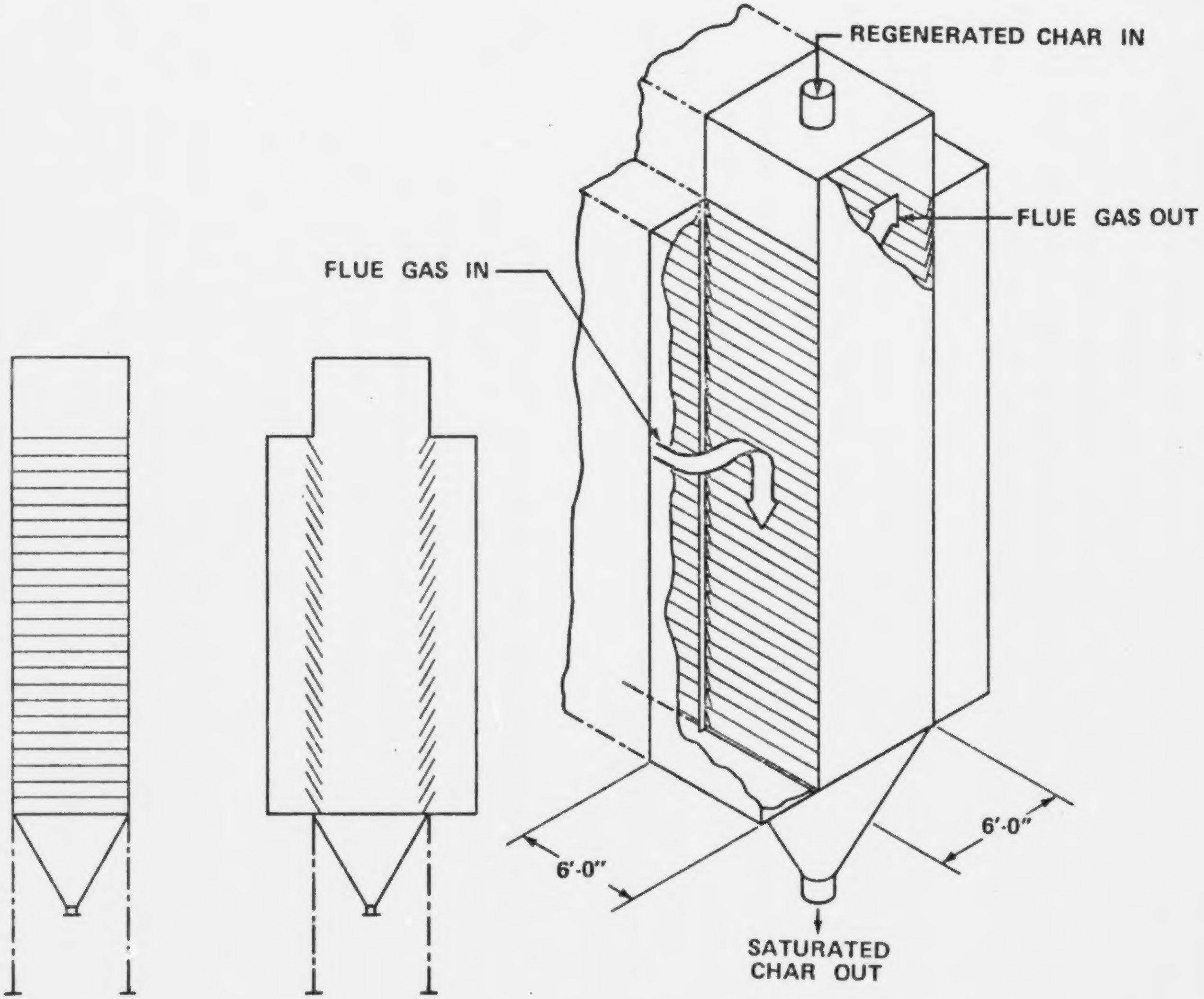


Figure 2.2.6-4. FW/BF Dry Adsorption Process Adsorber Module Detail.[243]

- the mixing action of fly ash with the char to prevent buildup at the inlet face of the char bed, and
- the minimization of entrainment of fines in the exit gas.

Design considerations for the char beds are the char depth, the number of modules placed in series to form a bed, and the number of beds in parallel. These criteria influence the degree of SO_2 removal, gas side pressure drop, and gas residence time. The char beds have been modularized for commercial application. The depth of the char bed is then set by the module choice. The number of modules in a bed and the number of parallel beds is determined by the SO_2 concentration in the flue gas and the removal efficiency desired. The modules in a bed are not separated by partitions and, therefore, are more conceptual than real. Consequently, the char can undergo horizontal movement (cross flow) with a possible increase in char attrition.[242]

The operating temperature of the absorber is primarily a function of the flue gas temperature and the temperature of the recycled char pellets from the regeneration section. The design operating temperature of the adsorber is approximately 280°F. Dilution air is used to cool the flue gas to this operating temperature. The char out of the regenerator must be adequately and uniformly cooled by a water quench prior to recycle to the adsorber.

Solids handling and control is another important consideration in the BF/FW adsorption section. Regenerated char must be uniformly distributed over an entire adsorber bed in a full-scale design. Solids handling equipment must be designed for the operating temperatures of the adsorption section.

Control of the adsorber is accomplished by adjusting the char flow rate and the amount of dilution air added to the flue gas. The char flow rate is not critical as long as enough char is fed to the adsorber to achieve the required degree of removal. Addition of dilution air should be carefully controlled to keep the flue gas temperature above the acid dew point. Also, a water quench might be an alternative cooling method for a commercial installation.

The regeneration reactor design is based on a Bergbau-Forschung unit that has been in operation over five years. The primary design factors are assurance of complete mixing, adequate residence time, and a noncombustible atmosphere. Conductive heat transfer using sand at about 1500°F as an inert heat transfer medium occurs to raise the char temperature to about 1200°F. The flow of the char and sand through the regenerator is gravity-induced. Char out of the adsorption section is transported by bucket elevator to a surge tank above the reactor to supply the necessary head to flow through the regeneration section. The fluid bed sand heater is likewise elevated to provide a head for gravity flow of the sand. The fluidized level is maintained by sand leaving through a refractory-lined overflow pipe which empties into the regenerator.

Complete and uniform mixing of the char and sand is very important. The mixing of two hot solids of widely differing particle size is difficult. Operating problems were reported at early pilot-scale installations, but BF has since improved the regenerator design.

No excess oxygen should be permitted to enter the regeneration vessel as this would increase the char consumption above that necessary for the chemistry of the regeneration process. For this reason, a reducing atmosphere with a slightly positive pressure is maintained in the vessel. Minimum resistance to upward flow of the evolved gases is also necessary. A means for the gas to move upward without having to find its way through the char-sand mixture may be required. Otherwise, evolution of gas may continue until after the char leaves the regenerator, resulting in "burps" in the char-sand separator.[244, 245]

The char cooler must quench the char pellets uniformly to prevent localized concentrations of hot char being recycled to the adsorber. The exit temperature of the char from the cooler should also be low enough so as not to initiate hot spots in the adsorber. The water quench of the hot char also must not produce thermal strains that result in breakage of the char pellets. BF feels that the char is strong enough to resist thermal stresses satisfactorily as long as air leakage into the unit is prevented.

Design considerations in the RESOX process for reduction of SO_2 are primarily concerned with the reaction of SO_2 gas with anthracite coal to produce elemental sulfur. Some factors that might influence this reaction are feed gas composition, pressure, temperature, residence time, gas distribution, and stoichiometry of carbon to SO_2.

Foster Wheeler reports that the pressure and gas distribution are not critical design considerations. The operating pressure is atmospheric. Proper gas distribution is accomplished by injection of the SO_2-rich gas at various ports around the lower vessel circumference. The stoichiometry of carbon to SO_2 is likewise not a sensitive parameter as long as an excess of carbon is present. Presently FW uses a carbon to SO_2 stoichiometry of about 2:1. About half of the RESOX coal feed leaves the bottom of the reactor as part of the RESOX ash stream. An economic incentive may exist to recover the heating value of the RESOX ash by burning it in the boiler.[246,247]

D. Applicability to industrial boilers[248,249]--The overall mechanical complexity of this system would appear to limit its applicability to large industrial boilers. In addition, the process employs rather large adsorbers and solids handling equipment, and unless extensive conveying systems are to be used, it will be necessary to locate both the adsorption and regeneration sections near the stack. The sulfur or sulfuric acid production unit can be located more remotely from the stack.

Another process characteristic that will limit the application of this process is that its long-term reliability is undetermined and may be a problem. In general, the process appears to be more applicable to large boilers than small boilers because of the large amount of regeneration equipment required. It does not appear cost effective for a small boiler application to use a complicated process requiring large amounts of equipment.

A potential process advantage is the ability to remove NO_x from the flue gas. Reported NO_x removal ability of the Bergbau system has varied from zero to near seventy percent. Recent data taken at the Scholz pilot unit has indicated that average NO_x removals of about twenty percent were being achieved.[250] At this time, however, the ability of the BF/FW process for NO_x removal has not been conclusively demonstrated.

Another attractive feature of this process is its ability to produce sulfur by using coal as the reductant. However, the process currently requires anthracite coal which has a limited availability. Current development efforts are underway by Foster Wheeler to evaluate methods of using bituminous coals as the reductant.

E. Summary--The Bergbau-Forschung/Foster Wheeler Process is a dry adsorption process which uses activated char pellets to adsorb SO_2 from flue gas. The char is thermally regenerated, and the SO_2 in the off gas is reduced in the presence of anthracite coal to elemental sulfur by a unit. The unique features of the process are the louvered wall design of the adsorber and the RESOX unit. The louvers influence the distribution and velocity of the flue gas and the char. RESOX is a proprietary process for the reduction of gaseous SO_2 to elemental sulfur. The major problems lie in the mechanical reliability of equipment and the availability of design and operating data for the RESOX unit.

2.2.6.4 Atomics International Aqueous Carbonate Process--

The Aqueous Carbonate Process (ACP) as developed by Atomics International uses an aqueous sodium carbonate (Na_2CO_3)solution to sorb sulfur dioxide (SO_2) from power plant flue gas. The dry scrubber product is treated to regenerate the scrubbing solution and to produce elemental sulfur. The technology for the regeneration and sulfur production steps is based principally upon established practice in the pulp and paper and the chemical industries.

A. System description--The ACP is divided into five major processing areas. These are gas cleaning, reduction, quenching and filtration, carbonation, and sulfur production. Figure 2.2.6-5 presents a simplified flow

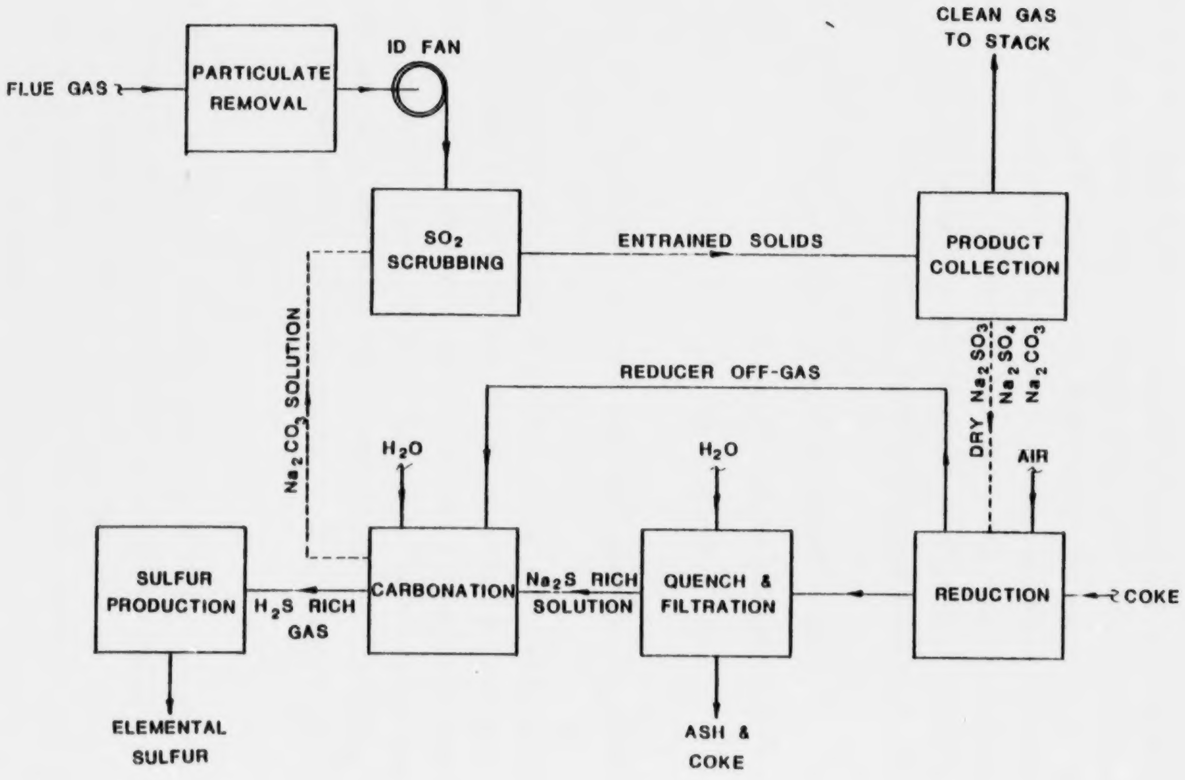


Figure 2.2.6-5. Simplified Flow Diagram for Atomics International Aqueous Carbonate Process.

diagram for this process illustrating the interaction of the various processing areas.

The SO_2 scrubbing and product collection equipment are combined in the gas cleaning subsystem. This subsystem employs a spray dryer for SO_2 scrubbing, and a bank of cyclones in series with an electrostatic precipitator for product collection and final particulate removal from the gas stream. Before the gas enters the scrubber, the ash content must be reduced to 1.0 gr/scf or less to limit the size of the ash removal equipment in the regeneration section. The following reactions occur in the scrubber.

$$SO_2 + Na_2CO_3 \rightarrow Na_2SO_3 + CO_2 \qquad (2.2.6\text{-}14)$$

$$Na_2SO_3 + \tfrac{1}{2}O_2 \rightarrow Na_2SO_4 \qquad (2.2.6\text{-}15)$$

$$SO_3 + Na_2CO_3 \rightarrow Na_2SO_4 + CO_2 \qquad (2.2.6\text{-}16)$$

Reaction 2.2.6-14 is the primary reaction for SO_2 removal.

Sulfur dioxide-clean flue gas exits the spray dryer and is routed to a bank of product collection cyclones, where the majority of the dry particles are removed. Final particulate removal is accomplished in high efficiency electrostatic precipitators or baghouses in which particulate emissions are limited to 0.01 grain/scf or less. Spent reactants from both the cyclones and precipitator or baghouse are collected and sent to the reduction section.[251]

The dry product collected in the gas cleaning system is stored in a silo and then conveyed pneumatically to the reducer vessel. This vessel contains a pool of molten salts at temperatures between 1700 and 1900°F. Carbon is injected in the form of petroleum coke (or coal). Combustion air is bubbled through the melt from injection nozzles in the vessel walls.[252]

In the molten salt pool, the following reactions take place:

$$Na_2SO_3(\ell) + 3/2C(s) \rightarrow Na_2S(\ell) + 3/2CO_2(g) \qquad (2.2.6\text{-}17)$$

$$Na_2SO_4(\ell) + 2C(s) \rightarrow Na_2S(\ell) + 2CO_2(g) \qquad (2.2.6\text{-}18)$$

$$C(s) + O_2(g) \rightarrow CO_2(g) \qquad (2.2.6\text{-}19)$$

Reactions 2.2.6-17 and 2.2.6-18 are endothermic. Reaction 2.2.6-19 is exothermic, however, and provides heat for both the endothermic reactions and system heat losses. The mechanism for reaction 2.2.6-19 is complex, involving sequential oxidation-reduction of the sulfur-containing salts as well as direct oxidation of carbon.[253,254]

The CO_2-rich off gas from the reducer is sent to the carbonation section after passing through a recuperator, waste heat boiler, and gas cooling tower. Reducer melt is continuously withdrawn and directed to the quench/dissolver vessel.

The sodium sulfide melt from the reducer is dispersed into fine droplets by steam shatter jets and dissolved in a green liquor solution which is near its boiling point. Insoluble material, mostly fly ash and reacted coke, is filtered out at this point using a rotary drum vacuum filter. Sodium is recovered from the ash filter cake using a simple washing technique. Both the quench and filtration operations are considered proven technology in the pulp and paper industry.

After the quench-filtration step, green liquor is contacted with CO_2-rich reducer off gas in a series of carbonation towers. The technology for this process step has been developed in the pulp and paper industry and proven processes are available. AI has developed their own carbonation scheme by modifying existing technology. The process reacts the CO_2-rich reducer off gas and green liquor from the ash filter, producing concentrated Na_2CO_3 solution for recycle to the gas cleaning subsystem plus an H_2S-rich Claus plant feed gas. A vent gas stream is also produced and sent to the Claus plant incinerator, and some additional ash is removed and sent to disposal.

The H_2S produced in the carbonation section is directed to a Claus plant for conversion to elemental sulfur. Claus technology is commercially available but has not been tested for specific application with this desulfurization process. The Claus plant tail gas and precarbonator off gas are combusted and returned to the spray dryer for treatment. AI proposes the use of an incinerator to burn these gases; however, they may be returned to the boiler for combustion.[255,256]

B. Development status--Development of the ACP has been underway since 1972. All of the process steps have been tested on a 1000 scfm scale or are considered proven technology by Atomics International. The spray dryer has been tested on pilot units at Bowen Engineering's North Branch, New Jersey facility and at Southern California Edison's Mohave Station. Early bench-scale work on the molten salt regeneration unit was performed at AI test laboratories in 1973. Based on this work, pilot scale tests were conducted with a feed rate equivalent to 1.25 MW. Bench-scale tests have also been conducted to determine the feasibility of using coal as the carbon source in the reducer. A fully integrated ACP system has never been operated, however.[257]

This process has been selected by EPA for demonstration at the 100 MW scale at Niagrara Mohawk Power Corporation's Charles R. Huntley Station. This will be the first full-scale demonstration of a regenerable process that can produce sulfur using coal as the reductant. A preliminary design and cost estimate have been completed and are under review. The current schedule for this process demonstration calls for plant startup to occur in late 1979 or early 1980.[258]

C. Design considerations[259,260,261]--Since this process uses a spray dryer for SO_2 absorption, the design considerations for the gas cleaning section of this process will be similar to those for the sodium based spray dryer process. Consequently, the reader is referred to Section 2.2.6.1 for a discussion of gas cleaning design considerations.

Sodium sulfite, sulfate, and unreacted carbonate are collected from the gas cleaning process area and sent to the spent absorbent storage hopper which has at least two days storage capacity. This is done in an attempt to decouple the operations of the gas cleaning and regeneration processing areas.

Operation of the molten salt reactor is the key design feature for this process. From a preliminary engineering viewpoint the design basis appears sound, as AI has identified potential problem areas associated with this reactor and have taken appropriate control measures. Historically, however, the operation of molten salt beds under reducing atmospheres has been particularly difficult to control, and it is likely that any unanticipated process problems would occur in this area.

Spent reactant, a source of carbon (petroleum coke or coal), and air are fed to the reducer to accomplish the reduction of sulfite and sulfate to sodium sulfide. Pilot plant data indicate that a steady-state reduction of 95 percent can be achieved with about 4 wt percent excess coke maintained in the reducer melt. Approximately one-half of the feed carbon is consumed chemically by the reduction reactions. The other one-half is burned to supply the heat needed for endothermic reduction to Na_2S, to melt the incoming salts, and to offset thermal losses and preserve the 1700 to 1900°F operating temperature. Typical reducer off gas contains less than 1 percent O_2 and 35 percent carbon oxides with a CO_2:CO ratio of 10 to 1 or higher. Heat is recovered from the off gas in a recuperator and waste heat boiler before the gas is cooled for use in the carbonation step.

The reducer can be continuously run at rated capacity or can be turned down to a standby condition corresponding to 5 percent of the design feed rate. To compensate for vessel heat losses and maintain operating temperature during standby operation, approximately 400 lb/hr of carbon is required in excess of that corresponding to the normal absorbent-to-carbon ratio. It will be important to avoid wide swings in melt temperature since appreciable spalling of refractory lining was noted in pilot tests under these conditions. For typical boiler load cycles or downtime, the preferred operating mode for the reducer is either 100 percent capacity or standby.

Because of its volatility, NaCℓ will be the predominant salt in the reducer off gas. It is likely that some of these salts will condense on the heat exchange surfaces of the recuperator and waste heat boiler and result in serious corrosion problems. AI has noted such deposits in pilot scale tests but claims, however, that the material can be easily removed with periodic washing or soot blowers. Most of the NaCℓ-rich particulates will be scrubbed from the gas in the gas cooling tower. A portion of the recirculating cooling tower liquid stream will be used as a chloride purge for the ACP.

The technology for quench and filtration of the reducer melt is considered standard practice in the pulp and paper industry. The design of equipment in this processing area borrows heavily from this technology.

The quench-dissolver is a thick-walled carbon steel vessel similar to those used in the paper industry. Some corrosion of the carbon steel is expected but the walls will be 5/8 inch thick to allow for it. Carbon steel is the preferred construction material for this application because it minimizes the possibility of embrittlement and cracking at the quench operating temperatures (around 200°F).

Two factors that are key to avoiding explosive conditions in the quench-dissolver are:

1) the melt must be broken up and dispersed into fine droplets, and
2) the green liquor should be maintained at or near its boiling temperature to avoid pressure excursions.

To assure operation in this mode, AI has specified that two completely redundant steam shatter systems be provided. One will use process steam generated in the waste heat boiler and the other will use low pressure steam generated in the boiler. Also, all pumps in the circulating liquor loop will be spared to insure an adequate water supply to the quench tank.

Filtration of fly ash and other solids from process liquor has caused problems in other alkali scrubbing systems. This situation is complicated in the ACP by the effect of high temperature reduction on the filtering characteristics of ash and unreacted coke in green liquor. Tests at AI's Molten Salt Test Reactor facility have shown that the fly ash and fuel ash present in the melt do not dissolve but do interact with other melt constituents to significantly alter their physical and chemical properties upon quenching. Laboratory tests have shown that it is possible to filter the ash and coke from simulated green liquor and concentrate the solids to a 40-60 wt percent solids filter cake. A prototype test loop has been prepared to demonstrate the rotary drum filter and other pieces of green liquor handling equipment.

Like the quench and filtration steps, the technology for carbonation is well developed by virtue of its use in the pulp and paper industry. Several commercial carbonation processes are available and can be applied to the ACP regeneration scheme. Most of these processes feature high energy and steam consumption, however, which has led AI to develop their own carbonation process. This process uses a series of sieve-tray columns to react the CO_2 from the reducer off gas with the sodium sulfide in the green liquor from the quench step. The diameters of the columns are set by the superficial velocity of the carbonation gas, in this case between 1 and 2 ft/sec. The heights of the columns are dictated by the mass transfer characteristics of the sieve trays. The key design criteria are H_2S and CO_2 uptake.

Production of elemental sulfur from the H_2S in the crystallizer off gas is standard Claus technology. Because of the low H_2S concentration (30-35 percent) of this stream, however, a three-stage Claus plant is required for 97 percent sulfur recovery. Other carbonation schemes produce a more concentrated feed stream and thus require a smaller Claus plant, but AI claims the overall economics favor their process concept.

D. Applicability to industrial boilers[262,263]--Overall chemical and mechanical complexity will limit the application of this system to small industrial boilers. As with the spray dryer throwaway system, space requirements may also limit its application; however, the regeneration section can be easily decoupled from the gas cleaning section and located in a peripheral plant access or completely off site.

One of the major process concerns of this system is operation of the high temperature reducing section. Since this section of the process has only been operated at the 1.25 MW scale, and then not in fully integrated operations, questions remain as to how the overall system will operate when coupled to a coal-fired boiler.

One of the main process advantages is that it provides a mechanism to produce sulfur without requiring a reducing gas. This advantage may not overcome the process disadvantage of increased complexity for applications to small industrial boilers.

E. Summary--The major advantage of the Aqueous Carbonate Process is its ability to produce sulfur with a solid carbonaceous reducing agent, eliminating the need for a reducing gas that must be produced from natural gas or by coal gasification. The decoupled nature of the process makes the gascleaning subsystem the sole operational interface with the flue gas source, and allows regeneration to be carried out at a distance from the scrubber. The need for reheat has been avoided in many applications by operating the spray dryer scrubber at L/G ratios too low to saturate the flue gas. A key to the success of the ACP is the successful operation of the molten salt reduction step which produces a very corrosive melt. Other potential problem areas of the process include the quench and solids filtration steps. Although the individual process steps have been demonstrated at the pilot plant level, integration of process components has not yet been demonstrated at any level of development.

2.2.6.5 Shell Flue Gas Desulfurization Process-

The Shell Flue Gas Desulfurization (SFGD) Process which is licensed by UOP's Air Correction Division is a dry metal oxide process. It is based on the ability of copper oxide to react with SO_2 in a flue gas and to be regenerated by hydrogen reducing gas. Although the Shell process is sometimes termed an adsorption process, the removal mechanism is the reaction of SO_2 with copper oxide to form copper sulfate according to Equation 2.2.6-20.

$$CuO + \tfrac{1}{2}O_2 + SO_2 \rightarrow CuSO_4 \qquad (2.2.6\text{-}20)$$

The process developers refer to this reaction as acceptance and to copper oxide as an acceptor. Copper is regenerated and SO_2 evolved by passing reducing gas, preferably hydrogen, over the copper sulfate. The end product of the process can be sulfur, SO_2, or sulfuric acid depending on the available markets and other economic considerations. Figure 2.2.6-6 is a simplified flow diagram of the system.

The unique feature of the process is a set of specially designed, parallel passage, fixed-bed reactors which use a copper oxide-on-alumina acceptor for SO_2 removal. The reactor configuration was specifically designed by Shell to maintain good catalyst stability and minimize pressure drop. The reactor cross section is proportional to the gas flow while the depth of the reactor determines the removal capacity. The SFGD Process can achieve 90 percent removal of SO_2 irrespective of inlet SO_2 concentration. The process also has demonstrated NO_x removal from the addition of NH_3 which reduces NO over CuO or $CuSO_4$ according to the following reaction.

$$2NH_3 + 2NO + \tfrac{1}{2}O_2 \rightarrow 2N_2 + 3H_2O \qquad (2.2.6\text{-}21)$$

A. Process description[264,265,266]--The system operates at about 750°F in the acceptance and regeneration steps. Flue gas is routed from a hot electrostatic precipitator through the Shell/UOP system before going to the air preheater.

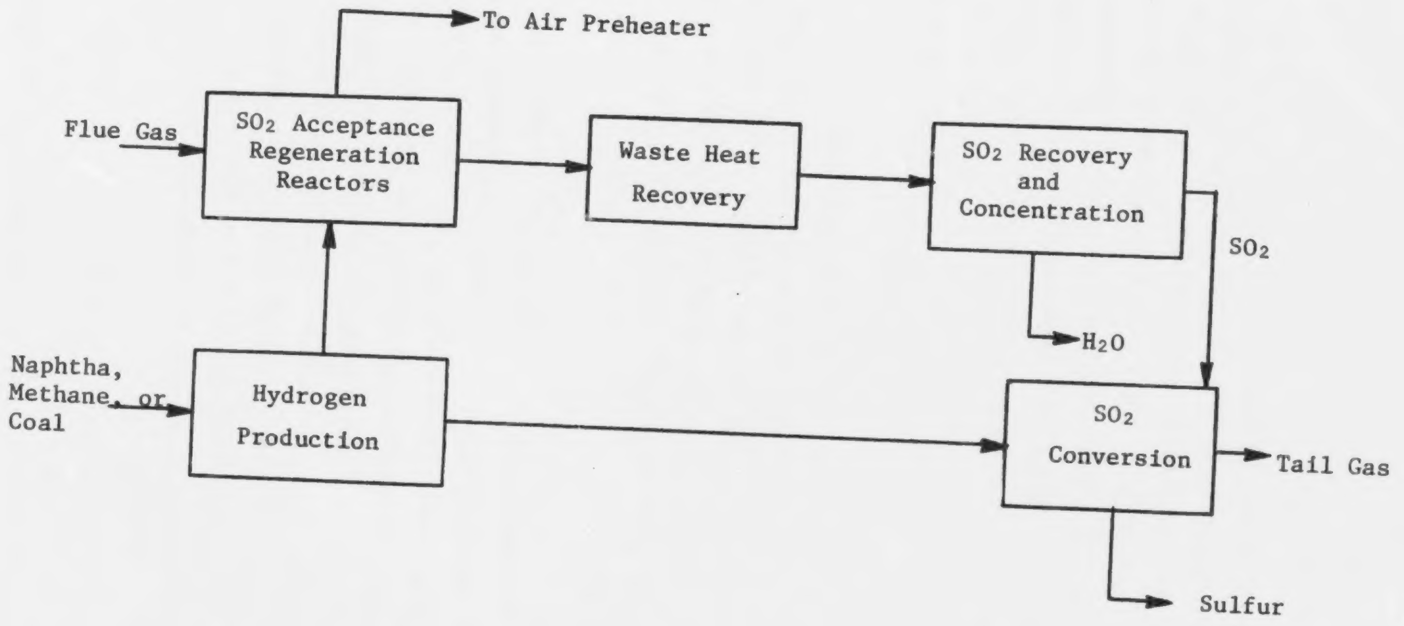


Figure 2.2.6-6. Block Flow Diagram - Shell Flue Gas Desulfurization Process.

The acceptance reaction depicted in Equation 2.2.6-20 and the regeneration reactions in Equations 2.2.6-22 and 2.2.6-23 are exothermic. Since acceptance and regeneration occur in the same temperature range, there is no energy lost in heating or cooling the beds.

When the majority of the copper in an acceptor has been converted to copper sulfate and SO_2 breakthrough begins to occur, the acceptor vessel is blocked off from the flue gas stream and purged in preparation for sorbent regeneration. The spent sorbent is regenerated by passing a diluted hydrogen containing reducing gas through the vessel. SO_2 is reevolved according to Equation 2.2.6-22. Swing reactor acceptors allow continuous flue gas treatment.

$$CuSO_4 + 2H_2 \rightarrow Cu + SO_2 + 2H_2O \qquad (2.2.6\text{-}22)$$

Additional hydrogen is consumed by the following side reactions. First, any CuO which was unused during SO_2 adsorption is reduced back to copper by Reaction 2.2.6-23.

$$CuO + H_2 \rightarrow Cu + H_2O \qquad (2.2.6\text{-}23)$$

After regeneration the acceptor is again purged and returned to flue gas service where any copper sulfide is reoxidized to copper sulfate by oxygen in the flue gas.

$$Cu_2S + \frac{5}{2}O_2 \rightarrow CuSO_4 + CuO \qquad (2.2.6\text{-}24)$$

Current SFGD designs use a combination of gas holder and a gas compressor for recovering the concentrating SO_2 from the regeneration gases. The gas holder is used to dampen flows from the cyclic absorption section to the gas compressor. The gas is compressed so that water can be condensed and separated from the recovered SO_2 before it is fed to the Claus unit.

The SO_2 conversion section presents several options. The concentrated SO_2 stream can be processed by Allied Chemical, RESOX, or BAMAG process technology, or a variation on one of these, to produce sulfur. However, sulfuric acid and liquid SO_2 production are options which appear, in general, to be more economical in terms of use of energy and reducing media, and are more proven technically.

Hydrogen is needed to regenerate Cu from the spent sorbent according to Reaction 2.2.6-22. UOP has generally proposed steam-methane or steam-naphtha reforming as a means of supplying needed reducing gas. These systems provide fairly high concentrations of H_2 compared to air-blown coal gasification which produces a gas highly diluted by nitrogen. A possible problem with using these systems is the decreasing availability of both methane and naphtha.

B. Development status--Shell has gone through several phases of testing but has yet to build a completely integrated unit applied to a coal-fired boiler. Extensive bench-scale testing to determine a suitable metal oxide acceptor, to select a reactor configuration, and to collect mass transfer data preceded the installation of a 400 to 600 scfm pilot unit at Shell's Pernis Refinery. The Pernis unit accumulated 20,000 hours of operating time during which various types of copper on alumina acceptors, regeneration agents, and reactor internal construction materials were examined. The flue gas used for the tests came from a high sulfur, heavy oil-fired heater. The stability of the acceptor when subjected to a flue gas environment was verified and the proposed reactor configuration was successfully demonstrated.

Information from Pernis was used to scale up and build a commercial-scale unit which went onstream in mid 1973 at the SYS refinery in Japan. The unit there consists of two acceptors operating on flue gas from an oil-fired boiler equivalent to about 40 MW of electric generating capacity

(75,000 scfm). Only the acceptors and water gathering SO_2 recovery system were installed and tested because refinery hydrogen and excess Claus plant capacity were available. The longest period of continuous operation has been about 2 months. The fairly successful SYS installation confirmed Shell's computer design methods for the reactor and their ability to scale up the system. Demonstration of both the viability of automatic sequencing controls and the large flue gas valves was an additional result.

The next step in Shell's development program was the testing of a reactor using coal-fired flue gas without particulate removal. This experiment took place at a Dutch utility station. The resistance of the reactor configuration to fouling by dust was examined and information was generated on erosion/corrosion tendencies of various materials of construction.[267]

Actual operation of the acceptance and regeneration steps in a coal-fired flue gas environment has been tested on a 0.6 MW unit at Tampa Electric's Big Bend Station. Results to date indicate no increase in pressure drop over time and stable catalyst activity. The system was operated with high fly ash loadings without noticeable effect on the acceptors.[268]
The other parts of the system, SO_2 concentration and sulfur production, are considered to be standard technology.

The availability of design information appears to be good but more confidence could be placed in the system's overall reliability if a fully integrated system were in operation. A fully integrated system could answer the troubling question of process operability under varying loads.

C. Design considerations[269,270]--Two major design considerations with the Shell/UOP process are: 1) the difficulty of integrating the cyclic behavior of the fixed-bed adsorption scheme with the variable SO_2 load from the power plant, and 2) the relatively inflexible operation of the hydrogen production facility and the Claus plant, if used. The fixed-bed acceptors themselves are not affected by the variations in SO_2 loading but the problem arises rather in maintaining efficient control of a widely varying situation.

Because unused copper oxide acceptor consumes H_2 during regeneration and because H_2 consumption is an important contributor to operating costs, acceptor operations should be optimized to minimize hydrogen consumption.

The capability of the H_2 plant to respond to frequently varying loads will be an important point, along with its economic viability and raw material supply, in deciding what type of hydrogen production facility to use. Also, sulfur producing plants such as Claus plants are rather inflexible at present. Two alternatives might be to produce sulfur from SO_2 using less expensive or more readily available reductants, or to instead produce sulfuric acid. Acid plants are more amenable to variable feeds, but good temperature control is still essential.

D. Applicability to industrial boilers--As with other regenerable systems, the application of this system to industrial boilers will be limited due to its overall complexity. In addition, this system requires hydrogen for the regeneration step. Hydrogen is either produced from raw materials whose future supply is uncertain (methane and naphtha) or from a coal gasification process which would add to the overall process complexity. In addition, space requirements for this process are more of a concern that with other processes due to the uncertain ability to decouple the acceptance and SO_2 recovery sections of the process.[271]

Retrofit applications will be limited due to the requirement of flue gas temperatures of about 750°F for the acceptance section of the process. Although there are no technical impediments to retrofitting this system it would require extensive duct work or reheating the flue gas to about 750°F.

E. Summary--The Shell Flue Gas Desulfurization Process has as its principal advantages the ability to remove NO_x simultaneously with SO_2, and the ability to produce a choice of three potential by-products: Sulfur, sulfuric acid, and liquid SO_2. The process is relatively complex, consumes hydrogen, and requires high temperature absorber operations. Process applications appear to be best suited to installations where NO_x removal must be achieved.

2.2.6.6 Chiyoda Thoroughbred 121 Process--

A. System description--The Chiyoda Thoroughbred 121 Process (CT-121) uses a jet bubbling reactor to absorb, oxidize, and neutralize SO_2 from flue gas. The reactor is a single vessel, consisting of flue gas inlet and outlet, air inlet, limestone slurry inlet, and gypsum slurry outlet. Air and mechanical agitation are also provided. A simplified flow diagram for this process is shown in Figure 2.2.6-7.

Hot flue gases are first routed through a prescrubber for gas cooling and particulate removal purposes. Chiyoda has reported that this prescrubbing step may be incorporated into the reactor.[272]

Flue gas is sparged into the reactor through an array of vertical spargers to generate a bubbling or froth layer. SO_2 is absorbed in the froth layer to form calcium sulfite which is oxidized to calcium sulfate. The froth layer provides good mass transfer for this SO_2 absorption reaction to occur. Cleaned flue gas is demisted and exhausted to the atmosphere.

Although the chemistry of this process is similar to that of conventional limestone scrubbing processes, it differs in that the sorbed SO_2 is oxidized to sulfate (gypsum), leaving only trace amounts of sulfite. Additionally, all chemical and process steps are carried out in one vessel.

The overall reaction for the system is described by the following equation:

$$SO_2 + CaCO_3 + \tfrac{1}{2}O_2 + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O + CO_2 \qquad (2.2.6\text{-}25)$$

Crystallized gypsum is discharged from the reactor as a slurry of 5 to 20 weight percent. Solids are separated and the liquor is returned to the reactor. The gypsum by-product is of a high quality and may be dewatered to produce a 90 percent dry product and landfilled. As an alternative,

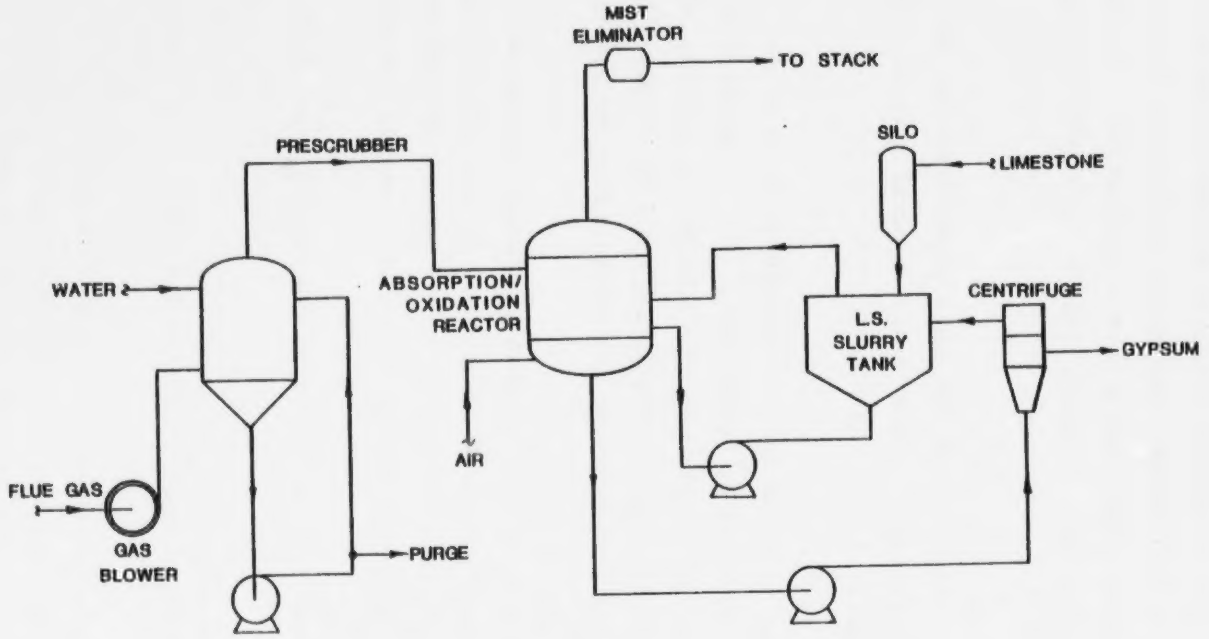


Figure 2.2.6-7. Simplified Flow Diagram for Chiyoda Thoroughbred 121 Process.

there is a possibility that the gypsum may be sold to the wallboard or portland cement industry.[273]

B. Development Status--Development of the CT-121 Process was initiated in 1975 in an effort to reduce the cost and complexity of the commercial CT-101 Process. Tests were initiated at the bench and laboratory scale and finally at the 650 scfm level. A conceptual design of a 125,000 scfm plant has been completed and plans are now under way to convert the 23 MW CT-101 system at Southern Company's Scholz Power Station to a CT-121 process configuration for further testing.[274]

C. Design Considerations--The scaling potential of this system is of concern in that calcium sulfate is formed in the reaction vessel. However, this system maintains a gypsum crystal concentration of 10 to 20 weight percent in the reactor which provides area for crystal growth. Consequently, gypsum should precipitate on the crystal surfaces and not form scale on the reactor walls.[275]

SO_2 removals of about 90 percent were achieved during the pilot plant testing. Higher removal efficiencies can probably be obtained, but would be at the cost of higher pressure drops and capital costs since it would be necessary to increase contact time by using larger scrubbers with more severe breakdown of bubbles by increased sparger pressure drop.

Control of the CT-121 process should be accomplished by monitoring pH levels in the absorber. The process should be rather easily controlled and rather insensitive to process upsets. However, ability of the process to follow load could be limited by sparger performance. As gas velocity through the sparger decreases, pressure drop decreases and bubble size increases as a result. This decreases mass transfer and SO_2 removal. It may be necessary to use multiple absorbers or a manifold system to distribute gas into any of a number of spargers in the absorber.[276]

D. Application to industrial boilers--Application of the CT-121 process to industrial boilers may be limited by space requirements for the scrubber and land area for product disposal. If the gypsum can be sold, by-product disposal problems would be eliminated. Marketability of gypsum will be extremely dependent upon local requirements for wallboard and, in general, it is expected that the product will have to be landfilled.

The process could potentially be decoupled, with the use of storage ponds for the slurry during centrifuge down time. This would disrupt normal process operations and increase land requirements. Ability of the process to follow load may be limited as discussed previously. Due to its relative mechanical and chemical simplicity, the process should be fairly acceptable for industrial boiler applications.[277]

E. Summary--The Chiyoda Thoroughbred 121 Process is a developing process that seems attractive for several reasons. Operation of the system should be relatively simple; capital costs and operating costs are potentially low. The process uses a plentiful raw material and produces a product that can be sold or landfilled. It is, however, undeveloped and there are many uncertainties as to its overall viability and costs. The upcoming demonstration of the plant at the Scholz Power Station will resolve many of these uncertainties.

2.3 CONTROLS FOR OIL-FIRED BOILERS

The major differences between oil- and coal-fired boilers are attributable to the differences in fuel properties. Since fuel oils are generally lower in sulfur and ash, their particulate and SO_2 emissions will be lower than for coal for a given sized boiler. The sulfur variability in fuel oil is also less than for coal which results in more stable FGD operations than required for coal-fired installations. In addition, it is possible to operate with as little as 5 to 7 percent excess air in oil-fired boilers whereas excess air requirements for coal-fired boilers may vary from 15 to 60 percent, depending on the method of firing. In general, these differences will not effect FGD process design considerations as discussed in this chapter, but will result in smaller FGD systems with correspondingly lower costs for the oil-fired FGD systems.

In the United States, there are currently 73 FGD systems being used to control SO_2 emissions from oil-fired boilers with an additional 19 systems under construction.[278] As shown in Table 2.3-1, all of these systems, except for one, are applied to oil-fired steam generating units which burn crude oil as a fuel. The non-oil field steam generating system is applied to an oil-fired boiler at a paper mill. Ninety-one of these FGD units are sodium throwaway systems and one is a double-alkali system. The actual designs of the systems are very similar to coal-fired FGD system designs except that the oil-fired systems have lower flue gas flow rates for a given boiler size because less excess air is used for combustion. Consequently the oil-fired FGD systems are smaller than coal-fired FGD systems would be for the same boiler size. Consequently, the discussions of FGD process design considerations in Section 2.2 for coal-fired boiler installations are applicable to oil-fired FGD systems.

TABLE 2.3-1. SUMMARY OF UNITED STATES OIL-FIRED INDUSTRIAL BOILER INSTALLATIONS[279]

Installation	Total Size (SCFM)	Number of FGD Units	SO_2 Removal Inlet (ppm)	SO_2 Removal Percent Removal	Type FGD
Alyeska Pipeline	50,000	1	150	96	Sodium
Belridge Oil	24,000	2	500	90	Sodium
Chevron	248,000	3	700	90	Sodium
Getty Oil	567,000	6	600	90-96	Sodium
Mead Paperboard	100,000	1	1500	95	Sodium
Mobil Oil	125,000	28	1500	90	Sodium
Texaco	380,000	32	1000	73	Sodium
*Belridge Oil	12,000	1	500	90	Sodium
*Chansler Oil	70,000	1	710	96	Double Alkali
*Getty Oil	396,000	4	600	90	Sodium
*Mobil Oil	80,000	7	500	85	Sodium
*Sun	12,000	3	700	85	Sodium
*Texaco	117,000	3	1000	95	Sodium
Total in Operation	1,544,000	73			
Total in Construction	687,000	19			
TOTAL -	2,231,000	92			

*Installations in construction phase

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SECTION 3

CANDIDATES FOR BEST SYSTEMS OF SO_2 EMISSION REDUCTION

In Section 2 of this report, eleven FGD processes were described and discussed with regard to their application to industrial boilers. These processes included both commercially available and developing technologies. In this section, these eleven processes are compared against each other in order to select those that appear to be the best candidates for industrial boiler application. The processes selected as best candidates will then be evaluated in detail with regard to their energy, environmental, and economic impacts. Results of the detailed evaluations will be used to compare the candidate processes for application to different sized boilers, with different coal types and SO_2 removal levels.

The processes selected as best candidates for application to industrial boilers are:

- Sodium Scrubbing
- Double Alkali
- Lime/Limestone
- Spray Drying
- Wellman-Lord

The first three processes are all currently used to control SO_2 emissions from industrial boilers throughout the United States. These processes are the ones of choice as evidenced by the fact that 118 of the 132 operating industrial boiler FGD systems are of these process types. The remaining operating industrial boiler FGD systems use ammonia process waste waters as a sorbent and are predominately found in sugar processing plants. Lime/

limestone, double alkali, and sodium scrubbing processes also appear to be the processes of choice for future installations as evidenced by the fact that 36 out of 39 systems in the planning or construction stages are of these process types.[1]

The spray drying process was selected as candidate technology for industrial boiler applications due to its potential for widespread use as evidenced by the large amount of interest expressed in this rapidly developing process. Presently, there are no full-scale spray drying FGD systems in operation; however, orders have been placed for five commercial spray drying systems, two of which are industrial boiler applications. Data on spray drying is available only from pilot scale units. This data will be used in later sections to compare the energy, environmental, and economic impacts of this process versus the other candidate processes.

The Wellman-Lord process was selected to compare the impacts of a regenerable FGD process against the other candidate processes which are all throwaway systems. There are currently no regenerable systems in operation on small industrial boilers, however, a demonstration of the Citrate Process is scheduled for operation in the near future. The Wellman-Lord process was selected as a candidate process over the Citrate Process primarily because of its increased development status and availability of data.

In order to select the candidate processes, a set of screening criteria was first developed that would insure consistency in evaluating all the processes. Next, a tabular comparison of the FGD processes was made for each screening criteria. Finally, the FGD processes judged to be the best candidates for industrial boiler applications were selected. The following sections describe the criteria that were developed for this study and present tabular comparisons of the FGD systems.

3.1 CRITERIA FOR SELECTION OF BEST SO_2 CONTROL SYSTEMS

As an aid to developing the screening criteria, several FGD process vendors and operators of industrial boiler FGD systems now in use or being planned were contacted to determine what factors they considered to be important in the application of FGD processes to their industrial boilers. The people contacted were asked why they selected the process they did; if they had been able to successfully operate it; and if they would use the same process if additional capacity were needed. Results of these contacts indicated that proven performance, process simplicity, and economics were the key factors for selecting a particular FGD process. Most FGD operators also indicated that they were satisfied with their process choices and seemed content with improving existing operations rather than switching processes.[2]

Evaluation or screening criteria were then established to provide an objective and consistent means of comparing the processes and to insure that the same factors were considered for each process. The screening criteria were then applied to each process and the results were compared and used to select the processes that appeared to be best suited for near-term industrial boiler applications. In some cases, engineering judgment was applied to rate the processes due to their relatively early development stage and lack of accompanying operating data. The screening criteria are listed in Table 3.1-1 and discussed in the following sections as they relate to the application of FGD systems to industrial boilers.

TABLE 3.1-1. FLUE GAS DESULFURIZATION SCREENING CRITERIA

1. Status of Development
   - Overall Process Development
   - Availability of Data

2. Performance
   - SO_2 Removal
   - Reliability

3. Applicability
   - Simplicity
   - Flexibility
   - Controllability
   - By-Product Marketability

4. Economic Considerations
   - Capital Investment Costs
   - Operating Costs

5. Energy Considerations
   - Liquid Pumping Requirements
   - System Pressure Drop
   - Regeneration Energy
   - Requirement for Reducing Gas

6. Environmental Considerations
   - Multipollutant Control
   - Secondary Pollutant Emissions

1) Development Status - These criteria considered both the overall process development and the availability of data. Development status is important because the more developed process will in general have more operating data and will provide a sounder data base for evaluating process economics.

2) Performance - Performance was interpreted to mean how well a system has demonstrated its desulfurization ability. Factors that are implicit in this definition are: SO_2 removal ability and system reliability. Particulate removal is often considered a performance criteria for FGD systems. However, for this process screening, particulate removal was assumed to occur upstream of the FGD process.

SO_2 removal ability is self-explanatory. Three removal efficiencies of 75, 85, and 90 percent were considered. Reliability is an important performance criteria since it can impact process capital and operating costs. Systems with poor reliability may require extra modules to maintain acceptable operations.

3) Applicability - The applicability of an FGD process to industrial boilers can be assessed by considering such factors as process simplicity, flexibility, controllability, and by-product marketability. Process simplicity is especially important to small industrial boiler applications, where FGD system capital and operating costs may become a signficant portion of the boiler expenses. In general, the less complex the process the more applicable it is.

Process flexibility was defined to be the ability to separate process steps within a system (decoupling). A process that can be easily decoupled can maintain operations in one section of the process while another section undergoes repairs. Process decoupling may also permit the concept of a centralized regeneration center. This concept may significantly reduce the overall costs of a regenerable system. However, for the concept to be viable, several FGD systems must be located in a close proximity.

Controllability refers to the relative ease of operating the FGD process and will have an effect on process operating costs. In general, the more easily controlled processes have lower operating costs which is a significant benefit particularly to small installations.

By-product marketability will also affect process applicability since available land-fill disposal space is often difficult to find. However, in some areas of the country, land-fill may be selected for economic reasons, whereas in other land-limited areas, it may be more economical to use processes which make saleable products such as gypsum, sulfuric acid, sulfur or liquified SO_2.

4) Economic Considerations - FGD process costs are very important for small industrial boiler applications and will probably be the major factor in selecting one FGD process over another. Both capital and annualized costs must be evaluated for each FGD system. For this preliminary screening, however, only rough process cost estimates could be made for the eleven FGD systems being evaluated. More detailed cost estimates of the selected candidate processes will be presented in Section 4.

5) Energy Considerations - Criteria to evaluate process energy requirements considered process features as liquid pumping requirements, system pressure drop, the relative amount of energy required for regeneration, and the requirement for reducing gas. Liquid pumping requirements and system pressure drops indicate the relative amounts of energy required for operating a system's pumps and fans. Regeneration energy requirements were based on an examination of the relative energy intensiveness of the unit operations used in each of the regenerable processes. The requirement for a reducing gas must be considered a process liability since reducing gases are produced either from reforming premium fuels such as naphtha or methane, or coal gasification. Either method of producing a reducing gas will add to process costs and complexity, in addition to increasing its energy consumption. Section 5 of this report will evaluate the energy usage of the candidate processes selected for industrial boiler application in detail.

6) Environmental Considerations - Criteria to evaluate environmental impacts considered two main items: 1) multipollutant control, and 2) secondary pollutant emissions. Multipollutant control capability is important since it has the potential of reducing the overall costs of an emission control system. Secondary pollutant emissions are important since it is undesirable for an FGD system to remove SO_2 at the expense of producing a secondary air, liquid or solid emission. Section 6 of this report will evaluate the environmental impact of the selected candidate processes in more detail.

3.2 SELECTION OF BEST CONTROL SYSTEMS

Each of the emission control techniques discussed in Section 2 can be designed to achieve any of the three emission control levels. The impact of increased SO_2 removal efficiency from 75 to 90 percent is predominately an economic one and will differ from system to system. In general, there will be less of an economic impact associated with the increased level of SO_2 control for the highly alkaline, clear liquor, sodium-based systems than there will be for the lime/limestone systems. This impact will be evaluated by performing economic sensitivity studies which will be presented in Section 4 for the selected candidate systems.

As discussed in Section 2, the impact of fuel type when going from coal- to oil-fired operations should not have an adverse affect on FGD systems. This fact is illustrated by several Japanese systems which have shown good operability on both coal- and oil-fired boilers. Consequently, this evaluation will consider only coal-fired operations.

All of the emission control techniques discussed in Section 2 will be considered in this section as possible candidates for the best control system for industrial boiler applications. The systems will be compared and discussed with regard to the criteria discussed in Section 3.1.

3.2.1 Development Status

One of the most important criterion for selection of candidate control systems was status of development. Table 3.2-1 shows the current status of development for the SO_2 control systems being considered. Of the operational processes, it is immediately apparent that the lime/limestone, double alkali, and sodium throwaway processes are the most highly developed and most widely used systems on industrial and utility boilers in the U.S. Wellman-Lord has recently undergone a demonstration on a coal-fired utility boiler with two additional utility units also under construction. The magnesium oxide process is currently operational at the 120 MW scale with three units in the planning stages.

With regard to systems in the planning or construction phases, the lime/limestone, double alkali, and sodium throwaway processes are again the predominate processes. The spray drying process has five systems in planning or construction phases which illustrates the large amount of interest in this rapidly developing technology. The remaining processes, Citrate, Bergbau Forschung/Foster Wheeler, Atomics International, Shell, and Chiyoda 121 are all currently in the demonstration stage and have no announced plans for commercial systems to be built in the near future.

3.2.2 Performance

Comments on SO_2 removal and reliability are highlighted for each of the candidate processes in Table 3.2-2. Early lime/limestone systems experienced many reliability problems, but better understanding of the process chemistry and improved process design has led to an improved reliability for these systems. Generally, clear liquor processes such as double alkali and sodium scrubbing have shown reliable operations. The first Wellman-Lord system, also a clear liquor system, applied to a coal-fired boiler in the U.S. has reported high system reliability during the first testing period. The magnesium oxide process has demonstrated greater than 90 percent SO_2 removal but has not been operated continuously for longer than eight days.

TABLE 3.2-1. OVERALL STATUS OF DEVELOPMENT OF CANDIDATES FOR BEST SYSTEMS OF SO_2 REDUCTION[3]

Process	Operational		Planned/Construction		Comments
	Industrial No.	Utility No.	Industrial No.	Utility No.	
Lime/Limestone	2	34	2	63	Total industrial capacity approximately 489,000 scfm. Total utility capacity approximately 43,000 MW_e.
Double Alkali	14	0	11	3	Total industrial capacity approximately 1,445,000 scfm. Total utility capacity approximately 1100 MW_e.
Wellman-Lord	0	2	0	4	Total utility capacity approximately 1850 MW_e.
Magnesium Oxide	0	1	0	3	Total utility capacity approximately 846 MW_e.
Sodium Scrubbing	102	3	23	1	Total industrial capacity approximately 5,081,000 scfm. Total utility capacity approximately 874 MW_e.
Spray Drying	0	0	2	3	Total industrial capacity approximately 73,000 scfm. Total utility capacity approximately 1300 MW_e.
Citrate/Phosphate	0	0	1	0	Total industrial capacity approximately 142,000 scfm.
Bergbau Forschung/ Foster Wheeler	0	0	0	0	No active units in U.S.
Atmocs International Aqueous Carbonate	0	0	0	1	100 MW_e demonstration unit planned for operation in 1980.
Shell/UOP	0	0	0	0	.6 MW_e pilot unit in operation
Chiyoda 121	0	0	0	1	Capacity approximately 23 MW_e.

TABLE 3.2-2. PERFORMANCE OF CANDIDATES FOR BEST SYSTEMS FOR SO_2 REDUCTION

Process	Comments
Lime/Limestone	Lime/limestone systems have demonstrated greater than 90 percent SO_2 removal.[4] In the past reliability has been a problem, but there is a trend toward better reliability.[5]
Double Alkali	Very high removals (>90 percent) and reliabilities have been demonstrated by double alkali systems on industrial boilers in the U.S.[6] High oxygen levels in the flue gas will lead to high sulfite oxidation, which results in sodium sulfate which is less active.
Wellman-Lord	The first application of the Wellman-Lord process to a coal-fired boiler in the U.S. has produced very good results. SO_2 removals greater than 90 percent are reported.[7] Reliability was high during the testing period. High levels of oxygen in the flue gas can promote sulfite oxidation which produces sodium sulfate which must be purged from the system.
Magnesium Oxide	SO_2 removal efficiencies of greater than 90 percent have been demonstrated.[8] Overall process reliability is difficult to evaluate since the longest continuous period of operation to date has been eight days.[9]
Sodium Scrubbing	SO_2 removal levels of greater than 90 percent are reported.[10] System reliability is generally excellent.
Spray Drying	Pilot unit test results have shown greater than 90 percent removal for low sulfur coal operations.[11] Reliability should be very good although only pilot scale units have been operated to date. A careful balance between flue gas condition and dryer operation is necessary to prevent condensation in downstream equipment.
Citrate/Phosphate	Pilot plant operations have reported up to 99 percent SO_2 removal.[12] Operation of the 64 MW_e unit at St. Joe Minerals will provide more information concerning performance and reliability.

TABLE 3.2-2. (Continued)

Process	Comments
Bergbau/Forschung/ Foster Wheeler	Operation of the BF/FW process in the U.S. at Gulf Power's Scholz Steam Plant experienced only limited success. Many mechanical problems were experienced which adversely affected process performance. The system at Lünen, Germany has reported much more successful operations.[13]
Atomics International Aqueous Carbonate	A fully-integrated Aqueous Carbonate Process system has never been operated. A 100 MW demonstration unit is planned for operation in 1980 which will help to answer performance and reliability questions.
Shell/UOP	Shell has yet to build a completely integrated unit applied to a coal-fired boiler. Questions about the system's performance and reliability remain.
Chiyoda 121	Conversion of the 23 MW_e CT-101 system at the Scholz Power Station to a CT-121 process is underway. SO_2 removals of about 90 percent were achieved during the pilot plant testing.[14] The process should be rather insensitive to flue gas variations, in fact, excess oxygen should be beneficial.

Three processes (double alkali, Wellman-Lord, and spray drying) are sensitive to flue gas conditions. Double alkali and Wellman-Lord are adversely affected by high oxygen levels in the flue gas since any oxidation products formed are less active thus requiring additional alkali and regeneration gases, respectively. Spray drying is sensitive to flue gas temperature, so the sorbent feed rate must be adjusted accordingly to avoid downstream condensation.

3.2.3 Applicability

Table 3.2-3 presents a summary of the applicability of the candidate FGD processes to industrial boilers. The throwaway processes (lime/limestone, double alkali, sodium scrubbing, spray drying and Chiyoda 121) are characterized by lower space requirements in the area near the boiler, fewer process steps, and no marketable by-product. These processes will, however, require land for solid waste disposal. The regenerable processes (Wellman-Lord, MgO, Citrate, Bergbau, Atomics International, and Shell/UOP) require more space near the boiler than the throwaway processes, have more process steps, and produce marketable by-products. In addition, they are generally more complex with sensitive controls. All the processes can be decoupled to some extent.

3.2.4 Economic Considerations

Preliminary cost estimates for 200×10^6 Btu/hr industrial sized FGD systems, which are reported in Table 3.2-4, were developed by directly scaling down utility FGD system cost estimates by process size. The utility costs were based on a series of cost estimates prepared by TVA for an Interagency Evaluation of FGD systems, and as such are developed on a consistent bases.[15]

TABLE 3.2-3. APPLICABILITY OF CANDIDATE SYSTEMS TO INDUSTRIAL BOILER

Process	Space requirements	Process simplicity	By-product
Lime/Limestone	Medium	Moderate	Sludge/Gypsum
Double Alkali	Medium	Moderate	Sludge/Gypsum
Wellman-Lord	High	Complex	Sulfur/Acid
Magnesium Oxide	High	Complex	Acid
Sodium Scrubbing	Low	Simple	High TDS Liquid
Spray Drying	Low	Simple	Dry Sodium/ Calcium Salts
Citrate/Phosphate	High	Complex	Sulfur
Bergbau-Forschung/ Foster Wheeler	High	Complex	Sulfur
AI Aqueous Carbonate	High	Complex	Sulfur
Shell/UOP	High	Complex	Sulfur/Acid
Chiyoda 121	Low	Simple	Sludge/Gypsum

TABLE 3.2-4. PRELIMINARY FGD SYSTEM COST ESTIMATES
(200×10^6 BTU/hr Sized Systems)[16]

Process	Preliminary capital cost (10^6 \$)	Relative Economic Rank
Lime/Limestone	1.9	1
Double Alkali	2.0	1
Wellman-Lord	2.3	2
Magnesium Oxide	2.1	2
Sodium Scrubbing	1.8	1
Spray Drying	1.8	1
Citrate/Phosphate	2.9	3
BF/FW	3.2	3
Atomics International-ACP	2.2	2
Shell/UOP	3.8	3
Chiyoda 121	*	1

*No capital cost data was available; however, the vendor claims capital costs should be about 60 percent of the capital costs of a limestone system.

Results of these estimates indicate that, in general, the mechanically simple throwaway processes will be less expensive than the more complex regenerable processes. This result is expected when one considers that users can be expected to choose an FGD process on the basis of cost, and essentially all the existing and planned FGD systems on U.S. utility and industrial boilers are of the throwaway type.

Of the throwaway processes, the sodium scrubbing and spray drying processes are shown to be the least expensive followed by the lime/limestone, and double alkali processes. As noted in Table 3.2-4 the cost for the Chiyoda 121 system was not available except for a statement by the vendor which claimed that its capital costs would be 60 percent of a limestone system.[17] If these costs do become verified upon further process development, the Chiyoda 121 process will become the least expensive.

Of the regenerable processes, the magnesium oxide, Wellman-Lord, and Atomics International process are shown to be the least expensive, with the Citrate, Bergbau, and Shell/UOP processes being somewhat more costly. The regenerable processes, except for the Wellman-Lord and magnesium oxide processes, are in relatively early development stages and consequently, their estimated costs are more speculative. A more in-depth cost analysis of the processes selected as being best suited for industrial boiler applications will be presented in Section 4 of this report.

3.2.5 Energy Considerations

A summary of the estimated energy requirements for the candidate systems is shown in Table 3.2-5. Results of these estimates indicate that the throwaway processes are less energy intensive than the regenerable processes. This is largely due to the energy required by the regenerable processes for regeneration of the SO_2 sorbent and for producing the by-product sulfur or sulfuric acid.

TABLE 3.2-5. FGD SYSTEM ENERGY REQUIREMENTS[18, 19]
(200 X 10^6 BTU/hr system, 3.5%S coal, 90% SO_2 removal)

Process	Liquid Pumping Requirements (L/G)	System ΔP	Regeneration Energy Required	Reducing Gas	Overall Energy Requirements (10^6 Btu/hr)*	Overall Energy Requirements Ranking
Lime/Limestone	High	Medium	No	No	6.4	Medium
Double Alkali	Low	Low	No	No	5.6	Low-Medium
Wellman-Lord	Low	Low	High	Yes	16.8	High
Magnesium Oxide	Medium	Medium	Medium	No	10.7	Medium-High
Sodium Scrubbing	Low	Low	No	No	5.6	Low-Medium
Spray Drying	Low	Low	No	No	2.6	Low
Citrate/Phosphate	Low	Medium	Low	Yes	11.8	Medium-High
Bergbau-Forschung/ Foster Wheeler	None	High	Medium	No	17.9	High
Atomics International Aqueous Carbonate	Low	Low	Low	No	8.7	Medium-High
Shell/UOP	None	Medium	Medium	Yes	9.2	Medium-High
Chiyoda-121	Medium	Medium	No	No	6.7	Medium

* Energy requirement estimates include energy for stack gas reheat of 3 X 10^6 BTU/hr except for the Spray Drying, Atomics International, Bergbau-Forschung, and Shell/UOP Processes.

This preliminary analysis indicates that the spray drying process will be the most energy efficient process, to be followed by the sodium scrubbing and double alkali processes, and the limestone and Chiyoda 121 processes. The low energy requirements of the spray drying system are due to the fact that no reheat energy is required. If reheat requirements were not included in these estimates, the overall energy required by the sodium, dual alkali, and spray drying processes would be essentially the same.

Of the regenerable processes the Atomics International, Shell/UOP, and magnesium oxide processes are the least energy intensive processes, and the Bergbau and Wellman-Lord processes are the highest energy users. Although the Wellman-Lord process has a relatively high rate of energy use, most of the energy consumed is in the form of low pressure steam which in many cases can be supplied from a waste steam source. A more detailed study of FGD process energy requirements is presented in Section 5 for the systems selected as best candidates for industrial boiler applications.

3.2.6 Environmental Considerations

The major environmental impacts considered for this process screening as shown in Table 3.2-6 were multipollutant control and secondary pollutant emissions. With regard to multipollutant control, all of the FGD processes under consideration have the potential for some particulate (fly ash) removal. However, removal of fly ash in the SO_2 scrubber of a regenerable system may lead to contamination of the by-product sulfur or acid and may also adversely affect the process chemistry. Consequently, particulate removal should occur upstream of regenerable systems. In general, the throwaway systems can remove both fly ash and SO_2 simultaneously in the gas-liquid contractor since a throwaway sludge is produced. Upstream particulate removal will probably, however, be required for the Chiyoda 121 process if a marketable quality product gypsum is desired.

TABLE 3.2-6. FGD SYSTEM ENVIRONMENTAL IMPACTS

Process	Multipollutant control		Secondary pollutants generated		
	Particulate removed	NO_x removal	Air	Solid	Liquid
Lime/Limestone	Yes	No	None	Sludge ($CaSO_3$, $CaSO_4$)	Possible leaching from waste solids
Double Alkali	Yes	No	None	Sludge ($CaSO_3$, $CaSO_4$)	Possible leaching from solids
Wellman-Lord	No	No	None	Relatively small Na_2SO_4 purge	Possible Chloride purge
Magnesium Oxide	No	No	Possible particulate emissions from calcining	None	Small purge stream
Sodium Scrubbing	Yes	No	None	None	High TDS wastewater stream
Spray Drying	Yes	No	None	Dry Sodium/ Calcium Salts	None
Citrate/Phosphate	No	No	Possible H_2S fugitive emissions	Relatively small purge Na_2SO_4	None anticipated
Bergbau-Forschung/	No	10-40%	Possible entrained carbon fines in exit gas	None	Small purge stream
Atomics International Aqueous Carbonate	No	No	Possible H_2S fugitive emissions	Fly ash filter cake	Small purge stream
Shell/UOP	No	Yes	Possible NH_3 emissions	Attrited copper oxide	Small purge stream
Chiyoda 121	No	No	None	Gypsum solids	Possible leaching from solids

In general, no appreciable NO_x removal will be experienced by the FGD systems except for the Shell/UOP process which has demonstrated up to 80 percent NO_x removal with ammonia addition. Up to 70 percent NO_x removal was also reported to occur on the carbon of the Bergbau process. However, additional testing has not confirmed the early results and this matter is still under investigation.[20]

As far as secondary pollutants are concerned, none of the processes should have significant air emissions. Significant quantities of solid wastes will be generated for the limestone, dual alkali, and spray drying processes. The limestone and dual alkali processes will produce a calcium sludge containing from 30-70 percent solids (rough calculations indicate a 58.6 MW_t(200 X 10^6 Btu/hr) system may produce about 2 tons per hour) and the spray drying process will produce a dry sodium or calcium salt waste material. Although the Chiyoda 121 process can produce a marketable quality gypsum, disposal of this material may be necessary if a market cannot be found. The amount of gypsum produced should be similar to the solids from the limestone process. The regenerable processes should not have major solid emission streams except perhaps for a Na_2SO_4 purge from the Wellman-Lord process if high oxidation occurs. No major liquid effluent streams are expected from these FGD processes except for the sodium scrubbing process whose disposal product is a high TDS aqueous stream that is composed mainly of dissolved sodium salts.

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SECTION 4

COST ANALYSIS OF CANDIDATES FOR BEST EMISSION CONTROL SYSTEM

Five flue gas desulfurization (FGD) processes were selected in Section 3 as being the best candidates for application to small industrial boilers. The processes selected were: Lime/Limestone, Sodium Scrubbing, Double-Alkali, Spray Drying, and Wellman-Lord. In this section the costs of applying each of these systems to the standard sized boilers will be considered. In addition, the cost impacts of achieving various SO_2 control levels, and of treating flue gases of fuels with different sulfur contents will be addressed. Costs presented in this section are estimates of installing a single FGD system on each of the model boilers.

In order to minimize the number of cases and yet cover the range of expected costs to be incurred, cost estimates will not be presented for the lime process. Lime process costs have been evaluated by TVA[1] and were found to generally be within 10 percent of limestone process costs. In addition, limestone is becoming the sorbent of choice for most new installations due primarily to the high calcining costs involved in producing lime from raw mined limestone.[2]

The general approach used in developing the process costs consisted of four main steps. First, a series of material and energy balance calculations were performed for each process, to establish process stream flow rates and energy requirements as functions of unit size, SO_2 removal, and the amount of sulfur in the coal. The energy requirements for each process are presented and discussed in Section 5 and the material balance results are presented in Appendix A. Second, each of the FGD processes were divided into a number of process areas, or modules, which represented separate processing areas. Third, equipment sizes were then developed for each process module based on the results of the material and energy balances. Finally, capital cost estimates were prepared by contacting process equipment vendors for price quotations in the size range for the standard industrial boilers of this study. All of the capital costs for each process area are reported in the form of direct capital costs which include all materials and labor installation costs. Except for the spray drying process, particulate control equipment costs were not included in this study.

In order to assess the reasonableness of the cost estimates calculated for this study, a series of graphs were prepared to illustrate the range of calculated installed costs versus site specific installed process costs reported in the EPA Industrial Boiler FGD Survey.[3] A range of calculated costs for a given process size is presented because of the cost differences attributable to coal sulfur content and SO_2 removal percentage.

Figure 4.0-1 presents the comparison of the limestone total capital costs, Figure 4.0-2 the dual alkali costs, and Figure 4.0-3 the sodium scrubbing costs. No comparison can be made for the Wellman-Lord process since there are no industrial boiler applications of this process in the United States.

There are currently two spray drying processes under construction on industrial boilers, but only one of those systems has reported costs published in the Industrial Boiler FGD Survey. That system treats 22,000 scfm of flue gas and has reported costs of $1.4 million. Spray drying costs calculated for this study on a 2.3 percent sulfur coal had estimated capital costs of $1.1 million for a 19,000 scfm system and $2.7 million for an 86,000 scfm system. Interpolating to the size of the reported spray drying system results in an estimated capital cost of $1.2 million which compares well with the reported cost for this process.

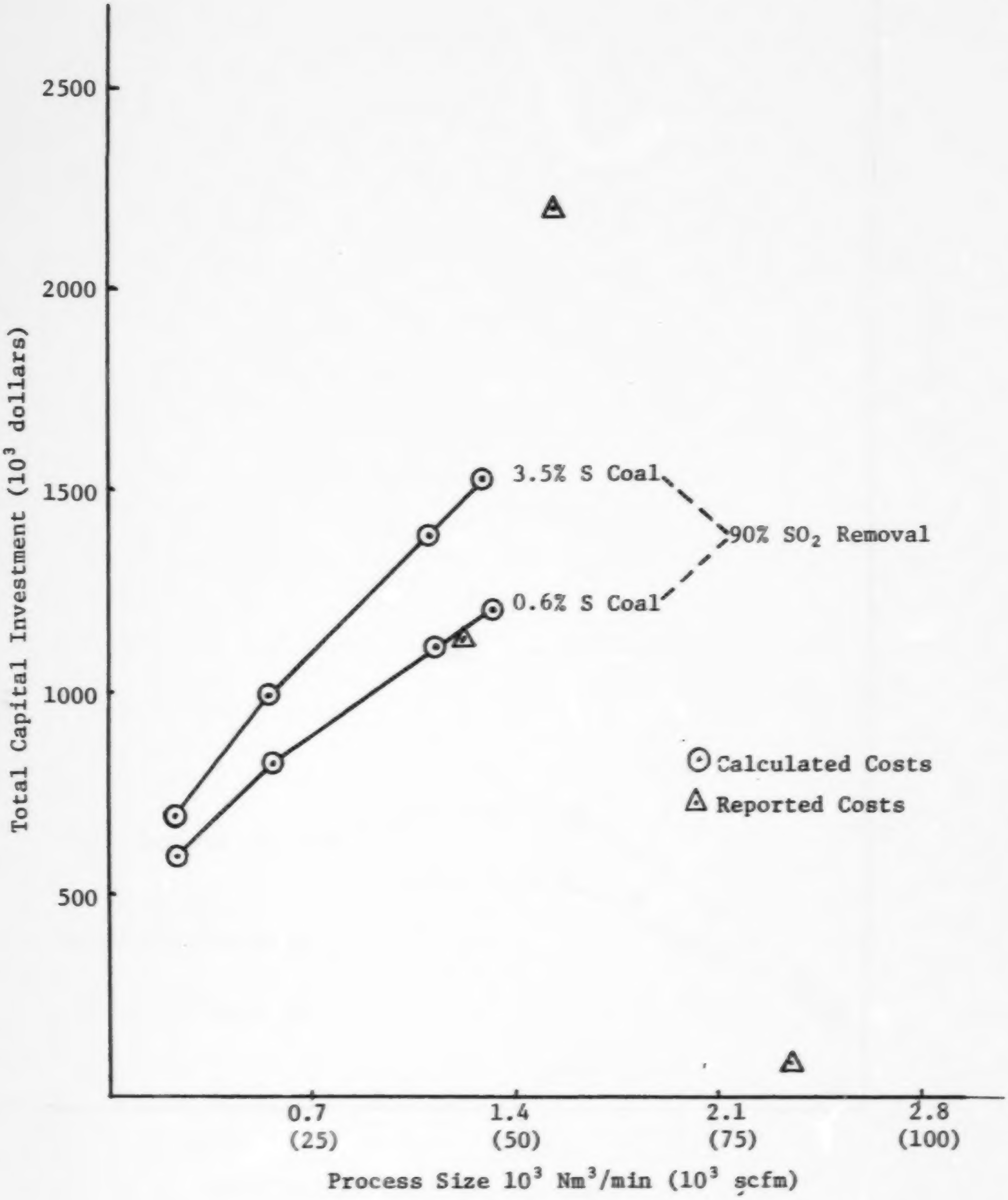

Figure 4.0-1. Comparison of calculated and reported total capital investment for the limestone process.

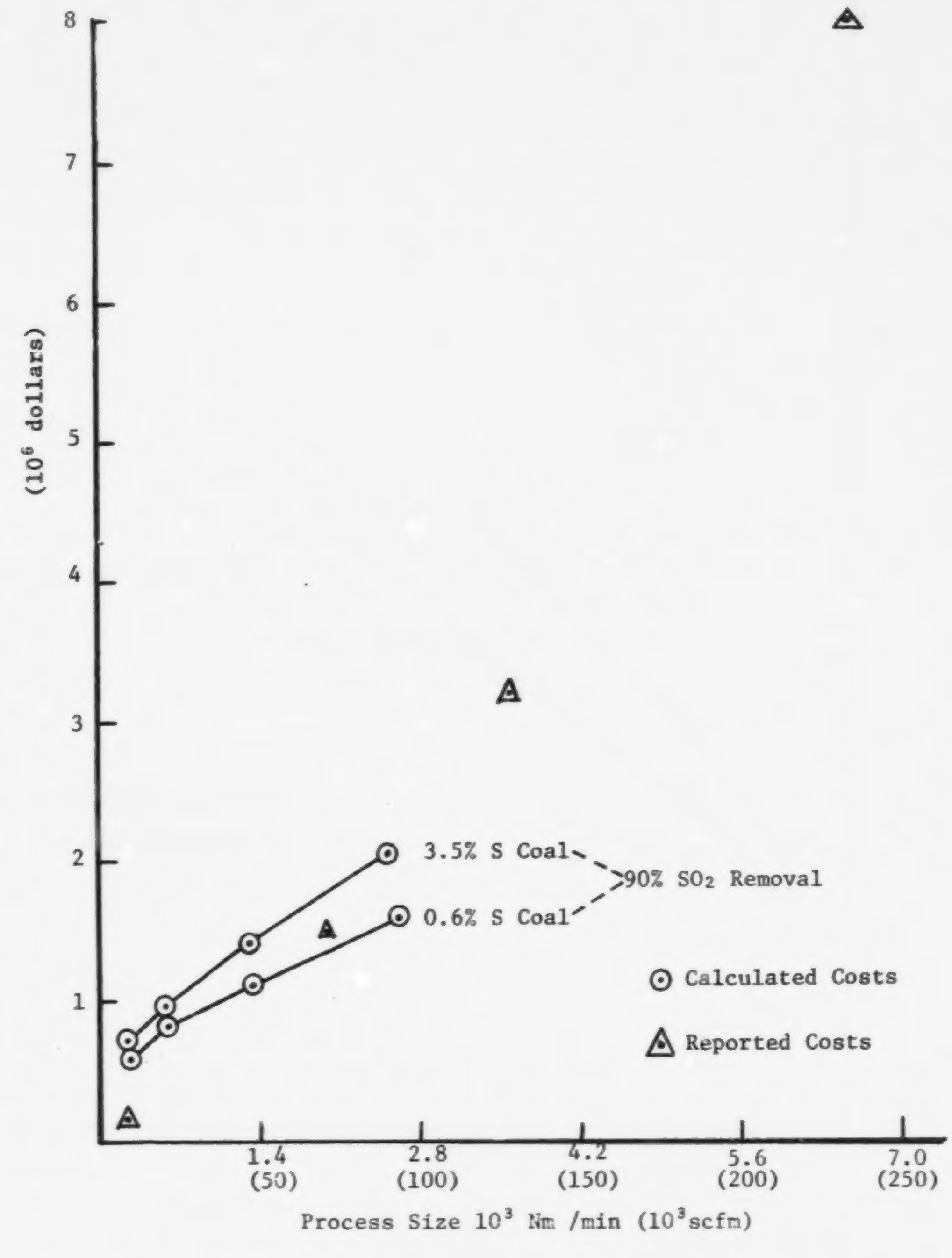

Figure 4.0-2. Comparison of calculated and reported total capital investment for the double-alkali process.

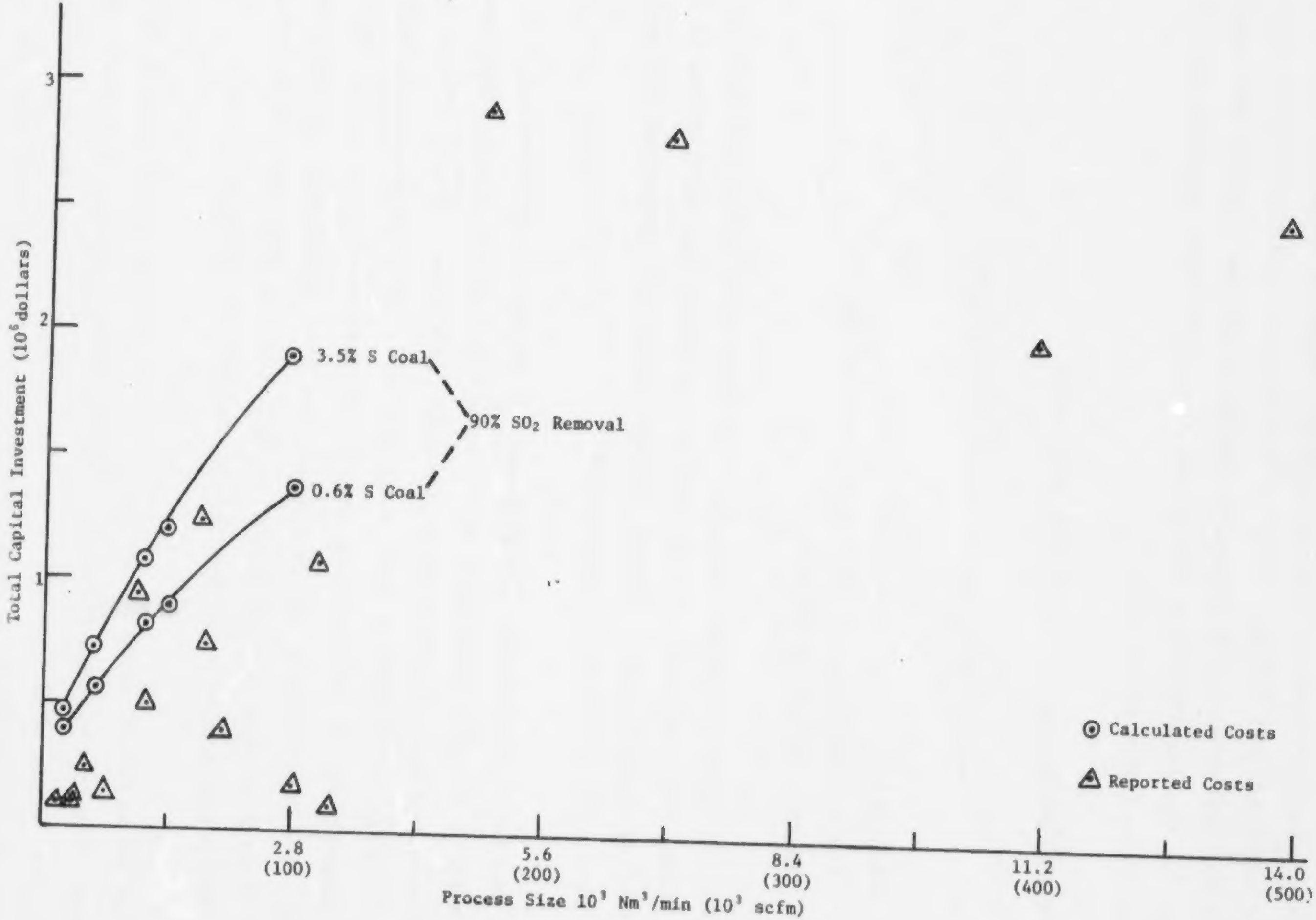

FIGURE 4.0-3. Comparison of Calculated and Reported Total Capital Investment for the Sodium Scrubbing Process (excludes wastewater treating costs).

The costs calculated for this study were for a general case and do not consider site specific factors that produce the range in reported costs shown in Figures 4.0-1 through 4.0-3. Although the process costs calculated for this study may not predict the actual installed costs for a given site specific case, they are representative of the actual installed costs reported by the process users as the Figures show.

4.1 CONTRIBUTORS TO CONTROL COSTS AND COST BASES

Annualized process costs consist of three components:

- annualized capital charges,
- direct operating and maintenance costs, and
- indirect operating costs.

The annualized capital charges component includes both direct and indirect costs. Direct costs consist of the cost of equipment and auxiliaries as well as the cost of installation. Direct capital costs were developed for each of the FGD processes in the following manner.

As mentioned previously, for purposes of estimating process capital costs, each of the five processes under consideration was divided into several process areas or modules which were individually costed. Their results were summed in order to develop overall process costs. This approach was used to permit comparisons of the costs of various process areas from one process to another. The major process areas defined for this evaluation and the process variables used to size the equipment in each process area are briefly described in the following sections as they impact the overall process costs.

The raw material handling and preparation process areas contain such equipment items as storage silos, conveyors, screw feeders, mix tanks, agitators, and pumps. This equipment is basically solid handling equipment and costs of the solids handling equipment were estimated to vary with the raw material feed rate for each process. Pumping costs varied with liquid flow rate through the process area. The raw material feed rates were calculated for each case and are presented in Appendix A. Feed rates were found to vary with boiler size, inlet SO_2, and SO_2 removal efficiency.

The SO_2 scrubbing and fans process areas are made up predominately of gas/liquid contacting and gas handling equipment. Accordingly, their costs were estimated to vary with the flue gas flow rate. Flue gas flow rates were calculated for each case and are presented in Appendix A. The major factor affecting flue gas flow is boiler capacity. In addition, the SO_2 scrubbing process area contained agitated tanks and pumps. Costs of these equipment items varied according to the liquid flow rate through the scrubber which was found to vary with boiler size, inlet SO_2 , and SO_2 removal efficiency.

The sizes of the regeneration, solids separation, purge treatment, and sulfur production process areas will vary with the amount of SO_2 being processed for each case. Accordingly, costs for these process areas were estimated to vary with the amount of SO_2 removed from the flue gas. The amount of SO_2 removed was calculated for each case and was found to vary with unit size, SO_2 inlet concentration, and SO_2 removal level.

The wastewater treatment processing area for the sodium scrubbing process consisting of oxidation and pH neutralization was assumed to be associated with the boiler or plant in question. Therefore, wastewater treatment appears as an operating cost only. Similarly, solids disposal for limestone and double alkali processes was assumed to be contracted out for offsite disposal and it, too, appears only as an operating cost.[5] The discharge rates were calculated for each of the cases and are presented in Appendix A. Unit size, SO_2 inlet concentration, and percent removal all affect the wastewater and sludge discharge rates.

Indirect capital costs are those not attributable to specific equipment items listed and include:

- engineering costs,
- construction and field expenses,
- contractors' fee,
- allowance for funds used during construction
- initial charge of chemicals and catalysts,
- start-up and performance test,
- contingency costs, and
- working capital.

Often the indirect cost items are estimated as a percentage of the direct capital costs or of the annual direct operating costs. Specific percentage used in this study are indicated in Table 4.1-1. Engineering costs were calculated only for the larger size processes and were assumed to be a constant value for the smaller process cases. This was done to reflect a more accurate impact of indirect costs for the small size FGD processes.

TABLE 4.1-1. INDIRECT CAPITAL COST FACTORS[6]

Item	Amount
Engineering	10% of installed cost of 58.6 MW_t ($200x10^6$ Btu/hr), eastern coal, 90% removal case for each process. Constant value for smaller process cases. 10% of installed cost of 118 MW_t($400x10^6$ Btu/hr), cases.
Construction and Field Expense	10% of installed cost
Contractor Fee	10% of installed cost
Start-up	2% of installed cost
Performance Tests	1% of installed cost or $2000
Contingency	20% of total direct and indirect costs
Land	0.084% of total turnkey costs
Working Capital	25% of total direct operating costs

Annualized capital charges can be calculated via several methods, including 1) utility financing, 2) private investor financing (discounted cash flow method), and 3) use of capital recovery factors. Capital charges can also include local taxes, insurance costs, and general and administrative (G&A) costs. By direction of EPA, the capital recovery method was used for this analysis. The capital recovery factor was calculated from Equation 4-1:

$$\frac{i(1+i)^n}{(1+i)^n-1} \qquad (4\text{-}1)$$

where: i = interest (10%)
n = number of year (15)

for this case the capital recovery factor is 0.13147. When G&A expenses of 4 percent are added, the annual capital recovery factor becomes 17 percent of the total capital investment.

Direct operating and maintenance (O&M) costs include:

utilities
raw materials,
operating labor,
maintenance and repairs,
fuel,
by-product credits, and
waste disposal.

Unit costs used to calculate the direct O&M costs are summarized in Table 4.1.2. Transportation costs for the raw materials were not considered in this study. The operating labor requirements were based upon one operator per shift and the supervision labor requirements were based upon 0.15 men/Shift.[7] Maintenance costs were based upon estimates reported by TVA.[8]

TABLE 4.1-2. VALUES USED FOR ANNUAL COST ITEMS[9]

Item		Cost
Direct Labor	\$/man-hour	12.02
Supervision	\$/man-hour	15.63
Maintenance Labor	\$/year	0.04(TDC)
Maintenance Material	\$/year	0.04(TDC)
Electricity	mills/kWh	25.8
Steam	\$/GJ	1.84
Process Water	$\$/m^3$	0.04
Methane	\$/GJ	2.05
Lime	\$/kg	0.0385
Limestone	\$/kg	0.0143
Soda Ash	\$/kg	0.0991
Solids Disposal[10]	\$/kg	0.044

Utility and fuel requirements were calculated for each case in a series of material and energy balances. Results of these calculations showed the variation in process energy requirements as a function of size, coal sulfur level, and SO_2 removal and are reported in detail in Section 5, Energy Impact of Candidates for Best Emission Control Systems. These calculations also serve as the basis for the energy costs reported in this section.

Results of the raw material and energy balance calculations also serve as the basis of the raw material cost estimates (see Appendix A). There are four basic raw materials used by the FGD processes under consideration: limestone, lime, sodium carbonate, and water. Other materials used by these systems such as steam and natural gas were costed according to energy equivalency. The amount of raw material used by each process was found to vary with both process size and the amount of SO_2 removed from the flue gas.

The third component of annualized costs is indirect operating costs. These costs include both payroll and plant overhead charges. In this study, payroll overhead was taken as 30 percent of the annual operating and supervision labor costs and plant overhead as 26 percent of the total annual labor and maintenance costs.[11]

4.2 CONTROL COSTS FOR COAL-FIRED BOILERS

Costs for each of the FGD processes being evaluated were estimated by the procedure discussed earlier in this section. Capital Costs for each FGD process area were estimated by obtaining equipment costs from the various equipment vendors and the wastewater treatment and solids disposal costs were supplied by PEDCo. Process stream flow rates and raw material requirements for the industrial boiler sized FGD systems were obtained from material balance calculations presented in Appendix A and were used to determine the detailed costs and serve as the basis of estimating the direct operating costs. Cost factors, as described in Section 4.1 were then applied to the direct capital costs in order to determine indirect capital costs and operating costs.

4.2.1 Interpretation of Results

Tables 4.2-1 through 4.2-5 summarize both the capital and annualized costs for each of the evaluated cases. These data are also plotted in a series of figures to be discussed later in order to illustrate the sensitivity of process costs to coal sulfur content, unit size, and percent SO_2 removal.

As shown in Tables 4.2-1 through 4.2-5, direct capital investment costs ranged from a low of \$187,000 for a sodium throwaway process (8.8 MW_t, 75 percent removal, 0.6 percent sulfur) to a high of \$2,573,000 for a Wellman-Lord process (58.6 MW_t, 90 percent removal, 3.5 percent sulfur coal). When indirect capital expenses are added, the total capital investment costs for these two cases become \$392,000 and \$4,233,000 respectively. From a capital cost standpoint, the sodium throwaway process appears to be the least costly process, and the Wellman-Lord the most costly process for all of the cases considered.

TABLE 4.2-1. LIMESTONE PROCESS COST SUMMARY

Boiler size and Type	Coal type	SO_2 removal level	Capital Costs		Annualized costs			% Increase over uncontrolled boilers
			Direct capital costs (10^3\$)	Total capital investment (10^3\$)	(10^3\$)	(\$/J/S)	(\$/MBtu/hr)	
8.8 MW_t (30 x 10^6Btu/hr) Underfeed Stoker	Eastern 3.5% S	90	358	691	441	0.051	14,700	41
		85	346	671	431	0.050	14,366	40
		75	332	646	416	0.049	13,866	39
	Western 0.6% S	90	305	594	362	0.041	12,066	33
		85	298	583	359	0.041	11,966	33
		75	289	569	354	0.040	11,800	32
22 MW_t (75x10^6Btu/hr) Chaingrate Stoker	Eastern 3.5% S	90	534	987	650	0.030	8,666	34
		75	490	913	593	0.027	7,906	31
	Western 0.6% S	90	448	821	458	0.021	6,106	22
		75	420	771	421	0.019	5,613	20
44 MW_t (150x10^6Btu/hr) Spreader Stoker	Eastern 3.5% S	90	760	1385	974	0.022	6,493	32
		75	696	1270	865	0.020	5,766	28
	Western 0.6% S	90	632	1108	576	0.013	3,840	18
		75	591	1041	537	0.012	3,580	17
58.6 MW_t (200 x 10^6Btu/hr) Pulverized Coal	Eastern 3.5% S	90	832	1530	1155	0.020	5,775	26
		85	797	1470	1102	0.018	5,510	25
		75	757	1392	1014	0.017	5,070	23
	Western 0.6% S	90	686	1201	634	0.011	3,170	14
		85	666	1168	614	0.010	3,070	14
		75	642	1125	586	0.010	2,930	13

TABLE 4.2-2. SODIUM THROWAWAY PROCESS COST SUMMARY

Boiler size and type	Coal type	SO_2 removal level	Capital Cost: Direct capital costs (10^3\$)	Capital Cost: Total capital investment (10^3\$)	Annualized costs (10^3\$)	Annualized costs (\$/J/S)	Annualized costs (\$/MBtu/hr)	% Increase over uncontrolled boilers
8.8 MW_t	Eastern	90	222	457	374	0.043	12,466	34
(30 x 10^6Btu/hr)	3.5% S	85	222	456	371	0.042	12,366	34
Underfeed Stoker		75	219	449	363	0.041	12,100	33
	Western	90	188	394	303	0.035	10,100	28
	0.6% S	85	187	392	302	0.035	10,066	28
		75	187	392	302	0.035	10,066	28
22 MW_t	Eastern	90	369	705	547	0.025	7,293	28
(75 x 10^6Btu/hr)	3.5% S	75	359	685	515	0.023	6,866	26
Chaingrate Stoker	Eastern	90	346	657	476	0.022	6,347	24
	2.3% S							
	Western	90	294	559	377	0.017	5,026	19
	0.6% S	75	291	554	374	0.017	4,986	19
44 MW_t	Eastern	90	553	1028	811	0.018	5,406	26
(150 x 10^6Btu/hr)	3.5% S	75	538	993	752	0.017	5,013	25
Spreader Stoker	Western	90	452	810	489	0.011	3,260	16
	0.6% S	75	447	800	479	0.011	3,193	16
58.6 MW_t	Eastern	90	615	1151	947	0.016	4,735	21
(200 x 10^6Btu/hr)	3.5% S	85	607	1129	902	0.015	4,510	20
Pulverized Coal		75	599	1108	865	0.015	4,325	19
	Western	90	499	886	529	0.009	2,645	12
	0.6% S	85	497	882	525	0.009	2,625	12
		75	494	875	519	0.009	2,595	12
118 MW_t	Eastern	90	937	1783	1532	0.013	3,830	19
(400 x 10^6Btu/hr)	3.5% S							
Pulverized Coal	Eastern	90	856	1588	1197	0.010	2,993	15
	2.3% S							
	Western	90	766	1355	760	0.006	1,900	10
	0.6% S							

TABLE 4.2-3. DOUBLE -ALKALI PROCESS COST SUMMARY

Boiler size and type	Coal type	SO_2 removal level	Capital Cost		Annualized costs			% Increase over uncontrolled boilers
			Direct capital costs (10^3\$)	Total capital investment (10^3\$)	(10^3\$)	(\$/J/S)	(\$/MBtu/hr)	
8.8 MW_t (30 x 10^6Btu/hr) Underfeed Stoker	Eastern 3.5% S	90	369	699	437	0.050	14,566	41
	Western 0.6% S	90	315	603	350	0.040	11,666	32
22 MW_t (75 x 10^6Btu/hr) Chaingrate Stoker	Eastern 3.5% S	90	522	960	625	0.028	8,333	32
	Eastern 2.3% S	90	489	889	519	0.024	6,920	28
	Western 0.6% S	90	442	799	425	0.019	5,666	20
58.6 MW_t (200 x 10^6Btu/hr) Pulverized Coal	Eastern 3.5% S	90	775	1,422	1,053	0.018	5,265	24
	Western 0.6% S	90	641	1,115	574	0.010	2,870	13
118 MW_t (400 x 10^6 Btu/hr) Pulverized Coal	Eastern 3.5% S	90	1,083	2,105	1,778	0.015	4,445	23
	Eastern 2.3% S	90	1,000	1,819	1,247	0.011	3,117	16
	Western 0.6% S	90	912	1,587	800	0.007	2,000	10

TABLE 4.2-4. SPRAY DRYING PROCESS COST SUMMARY

Boiler size and type	Coal type	Sorbent type	SO_2 Removal level	Capital Cost		Annualized costs			% Increase over un-controlled boilers
				Direct capital costs (10^3 \$)	Total capital investment (10^3 \$)	(10^3 \$)	(\$/J/S)	(\$/MBtu/hr)	
17.6 MW_t (60 x 10^6 Btu/hr) Chaingrate Stoker	Western 0.6% S	Lime	75	450	828	432	0.025	7200	
22 MW_t (75 x 10^6 Btu/hr) Chaingrate Stoker	Eastern 2.3% S	Lime	75	582	1035	520	0.024	6933	28
44 MW_t (150 x 10^6 Btu/hr) Spreader Stoker	Western 0.6% S	Sodium	90	845	1450	695	0.016	4633	22
			75	834	1420	644	0.015	4293	21
			50	826	1401	607	0.014	4047	19
		Lime	90	865	1460	630	0.014	4200	20
			75	850	1431	592	0.013	3947	19
			50	839	1408	565	0.013	3767	18
58.6 MW_t (200 x 10^6 Btu/hr Pulverized Coal	Western 0.6% S	Sodium	75	917	1555	718	0.012	3590	16
118 MW_t (400 x 10^6 Btu/hr) Pulverized Coal	Eastern 2.3% S	Lime	70	1573	2682	1242	0.011	3105	16
	Western 0.6% S	Lime	70	1501	2503	946	0.008	2365	12

TABLE 4.2-5. WELLMAN-LORD PROCESS COST SUMMARY

Boiler size and type	Coal type	SO removal level	Direct capital costs (10^3\$)	Total capital investment (10^3\$)	Annualized costs (10^3\$)	Annualized costs (S/J/S)	Annualized costs (\$/MBtu/hr)	% Increase over uncontrolled boilers
8.8 MW_t (30 x 10^6Btu/hr) Underfeed Stoker	Eastern 3.5% S	90	796	1539	558	0.063	18,600	52
	Western 0.6% S	90	370	896	385	0.044	12,833	35
22 MW_t (75 x 10^6Btu/hr) Chaingrate Stoker	Eastern 3.5% S	90	1420	2483	809	0.037	10,786	41
	Western 0.6% S	90	732	1440	529	0.024	7,053	24
58.6 MW_t (200 x 10^6Btu/hr) Pulverized Coal	Eastern 3.5% S	90	2573	4233	1289	0.022	6,445	29
	Western 0.6% S	90	1354	2378	771	0.013	3,855	17

With regard to annualized costs, the relative rankings of the FGD processes remain the same as with the capital costs for all cases considered. The sodium throwaway process again emerged as the least costly alternative. It should be noted, however, that part of the low costs for this process are attributable to the relatively simple water treating process consisting of oxidation plus pH neutralization. If a more elaborate water treating scheme were required, possibly to comply with zero discharge regulations, process costs would increase accordingly and could even make this process the most costly alternative.

Figures 4.2-1 and 4.2-2 illustrate this graphically by showing the variation in total capital investment costs and total annualized costs with unit size for each process applied to high sulfur coal at 90% removal. Figure 4.2-1 indicates a capital cost increase of 100-200 percent in each of the processes in going from an 8.8 MW_t ($30x10^6$ Btu/hr) to a 58.6 MW_t ($200x10^6$ Btu/hr)size. Figure 4.2-2 shows the annualized costs for the processes as a function of unit size. As with the capital costs, almost a 200 percent increase in annualized costs will occur in going from 8.8 MW_t to a 58.6 MW_t size. These figures indicate that the relative annual cost of the Wellman-Lord process is less than its capital costs when compared to the other FGD processes. This is due to the credits received for the production of sulfur as a by product.

Figures 4.2-3 and 4.2-4 show the variation in total capital costs and total annualized costs with unit size for the candidate processes applied to low sulfur coal boilers at 75% removal. One can see from these plots that the sodium throwaway process again has both the lowest capital and annualized costs for low sulfur coals requiring low SO_2 removal levels. It should be noted here that the capital costs of the spray drying systems include costs for a baghouse as a particulate collection device which is an integral part of the spray drying system. The annual costs of the spray drying systems also include the

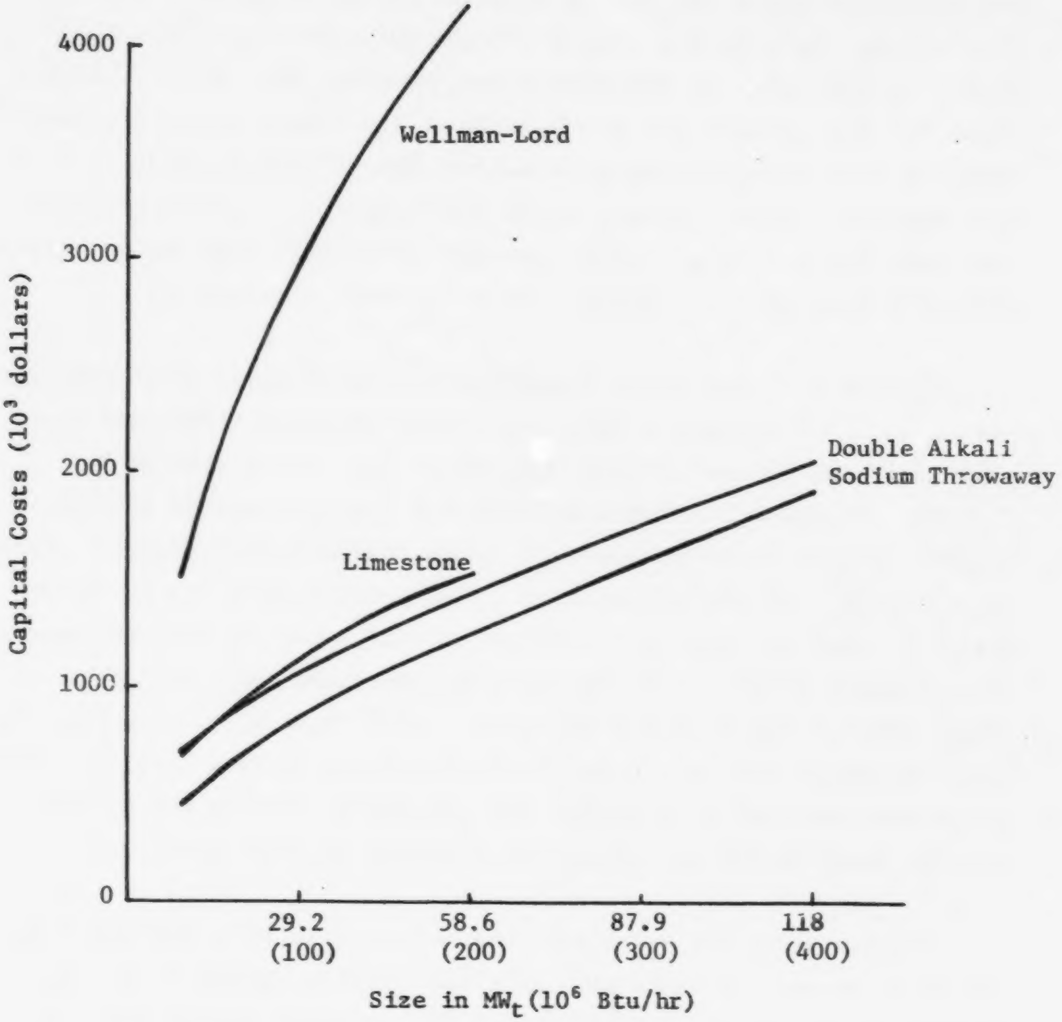

Figure 4.2-1. FGD capital costs *versus* unit size. (3.5% S coal, 90% removal)

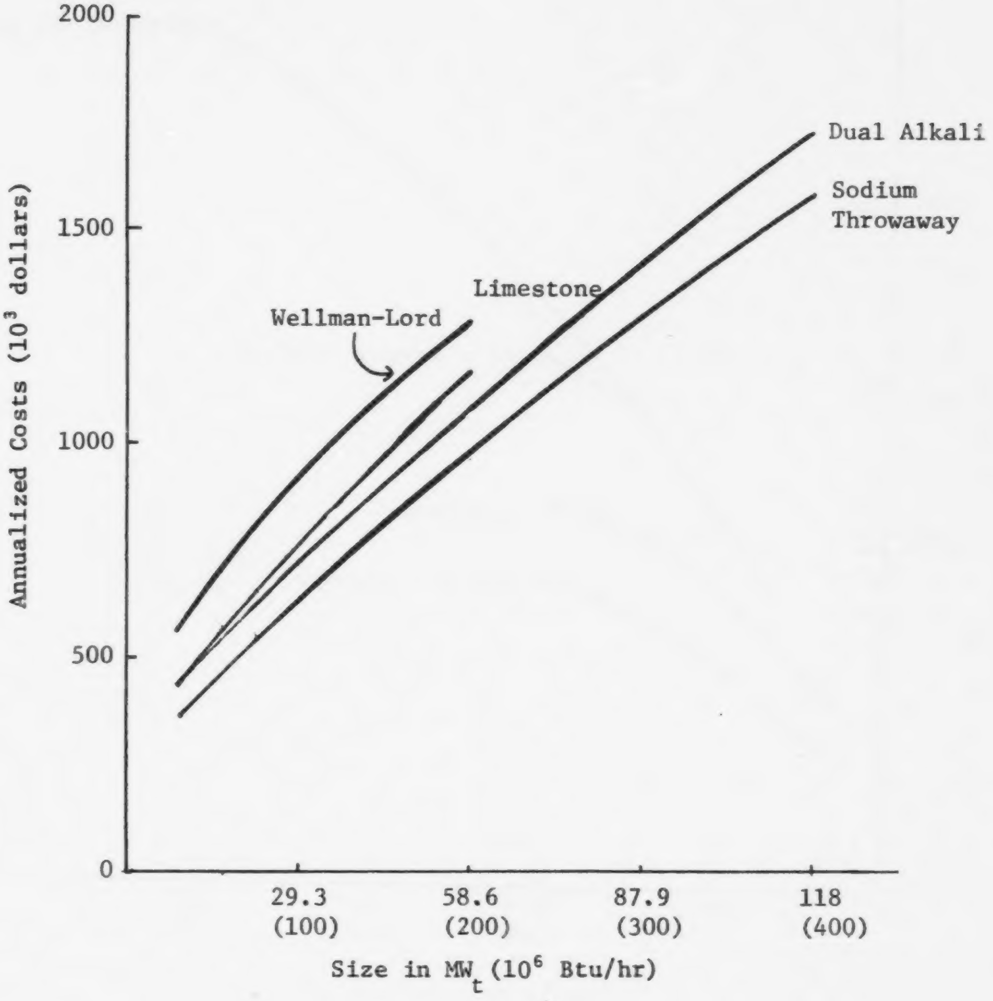

Figure 4.2-2. FGD annualized costs *versus* unit size. (3.5% S coal, 90% removal)

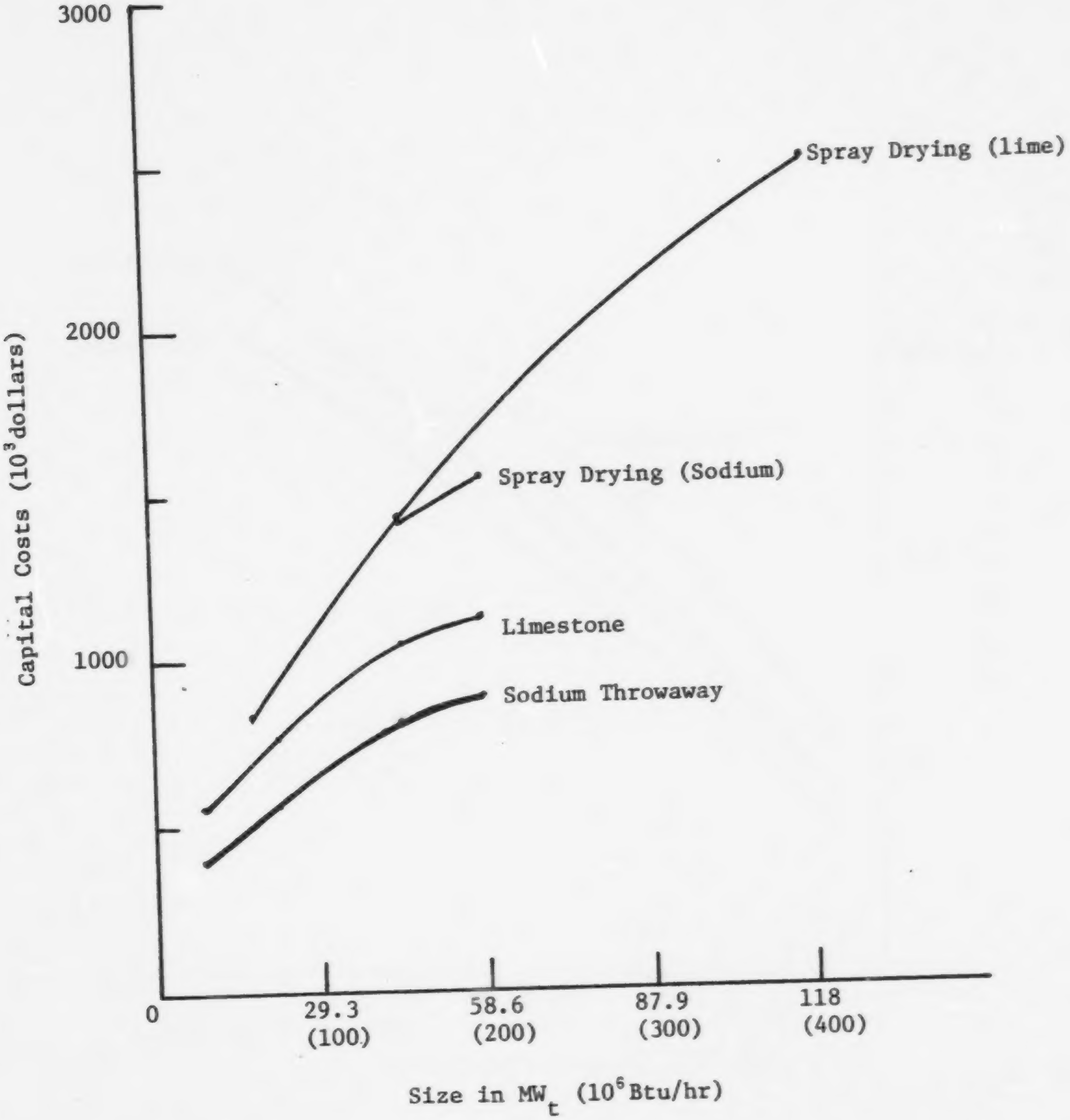

Figure 4.2-3. FGD capital costs *versus* unit size. (0.6% S coal, 75% removal)

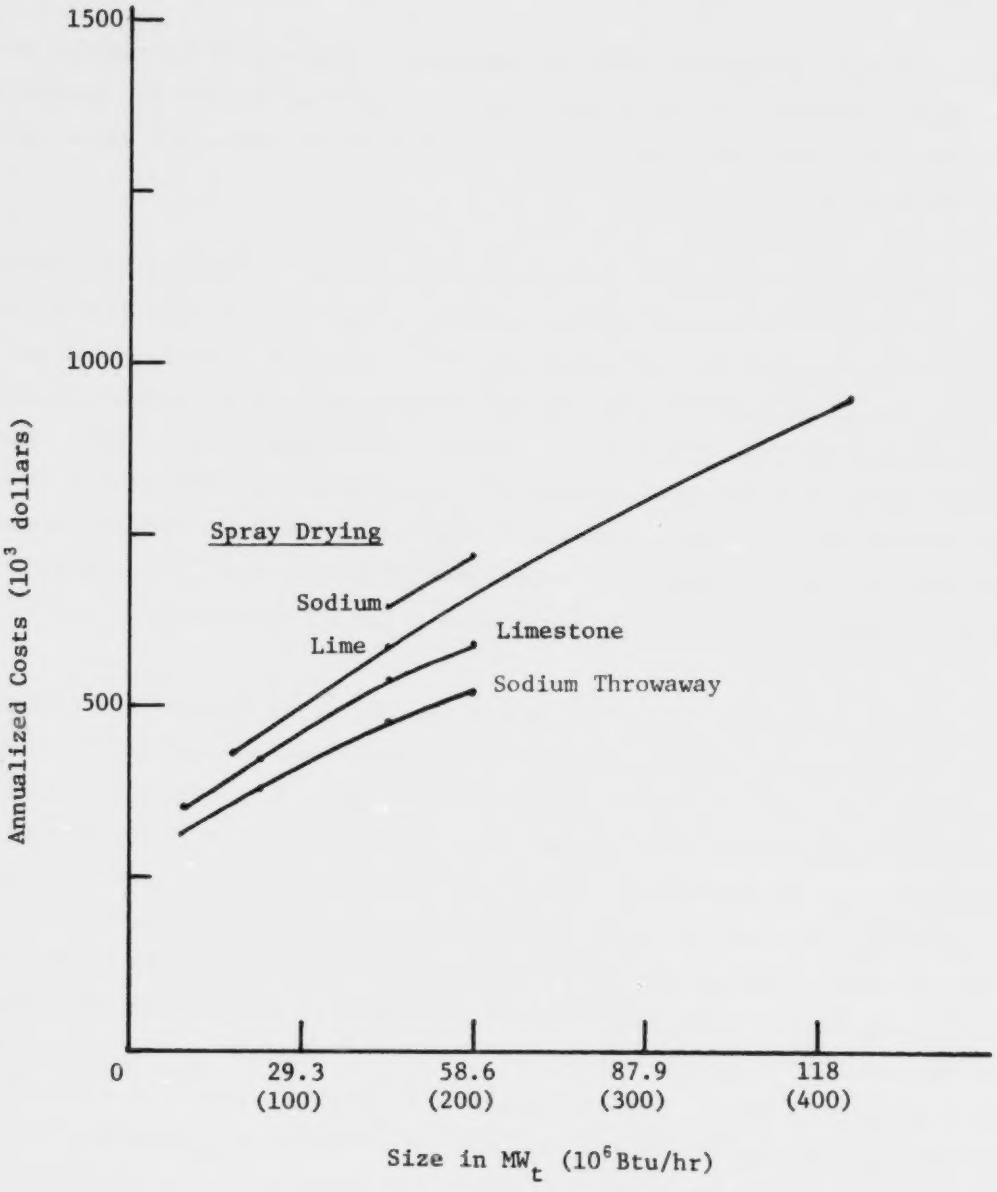

Figure 4.2-4. FGD annualized costs *versus* unit size. (0.6% S coal, 75% removal)

capital charges associated with the baghouse. However, to be consistent with the other candidate processes which were designed solely for SO_2 control, disposal costs were only charged for that portion of the solid waste stream associated with SO_2 removal.

Figures 4.2-5 and 4.2-6 illustrate the changes in capital and annualized costs as a function of coal sulfur content. Figure 4.2-5 illustrates the capital cost impacts and indicates that for a given FGD process size and SO_2 removal level, the capital cost of the FGD processes will increase about 50 percent in going from a 0.6 to 3.5 percent sulfur coal. Also, with regard to capital costs, the relative ranking of the processes did not change with an increase in the coal sulfur content. Figure 4.2-6 illustrates the annualized cost impacts and indicates that annualized costs for the FGD processes will increase almost 100 percent in going from a 0.6 to a 3.5 percent sulfur coal.

Figures 4.2-7 and 4.2-8 illustrate the change in capital and annualized costs as a function of SO_2 removal for the limestone and sodium throwaway processes applied to the 58.6 MW_t boiler burning 3.5 percent sulfur coal. Since the double-alkali and Wellman-Lord processes were only considered for the 90 percent removal cases, their cost variation with SO_2 removal could not be assessed. Capital costs of the limestone and sodium processes are shown to increase about 5-10 percent in going from an SO_2 removal of 75 to 90 percent. This low variation is due to the fact that no flue gas by-pass was assumed for the calculations, and consequently, the gas handling costs remain constant for a given unit size. Additional cost sensitivity cases will be developed for flue gas by-pass operations and will be reported in a separate section of this report. Figure 4.2-8 shows that the annualized costs for both these processes increase about 10 percent in going from 75 to 90 percent removal.

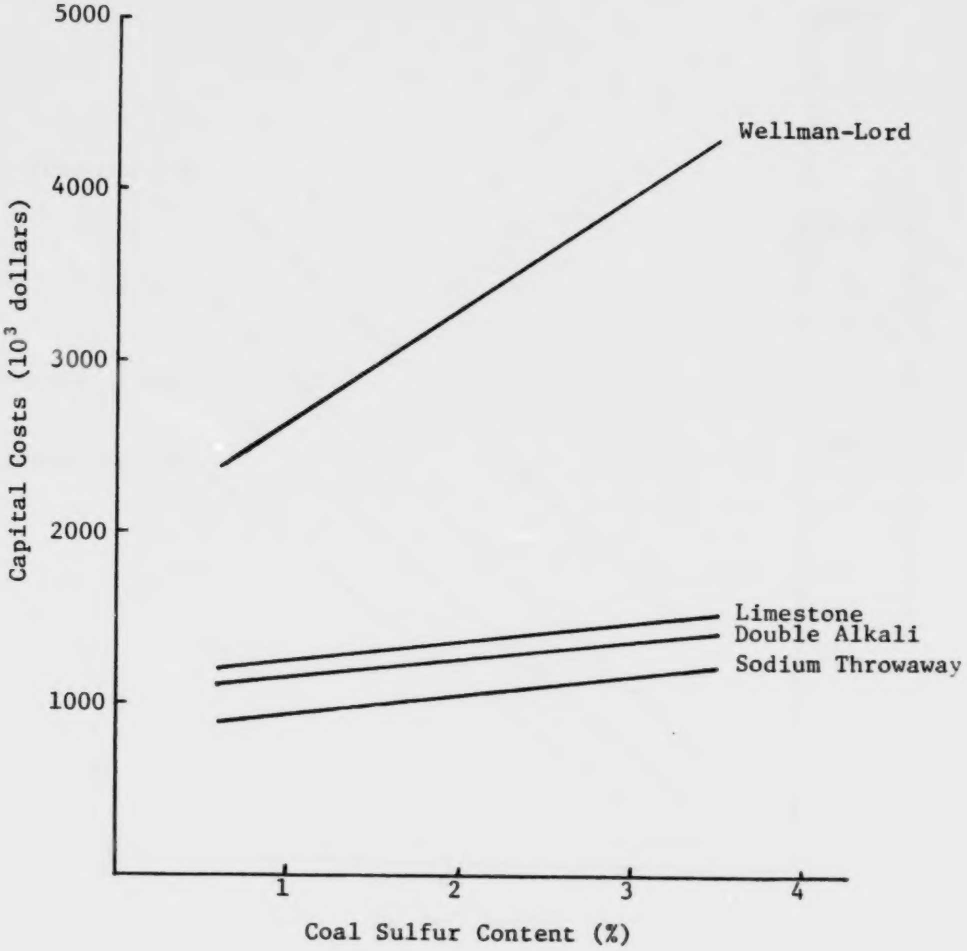

Figure 4.2-5. FGD capital costs *versus* coal sulfur content. (58.6 MW, 90% Removal)

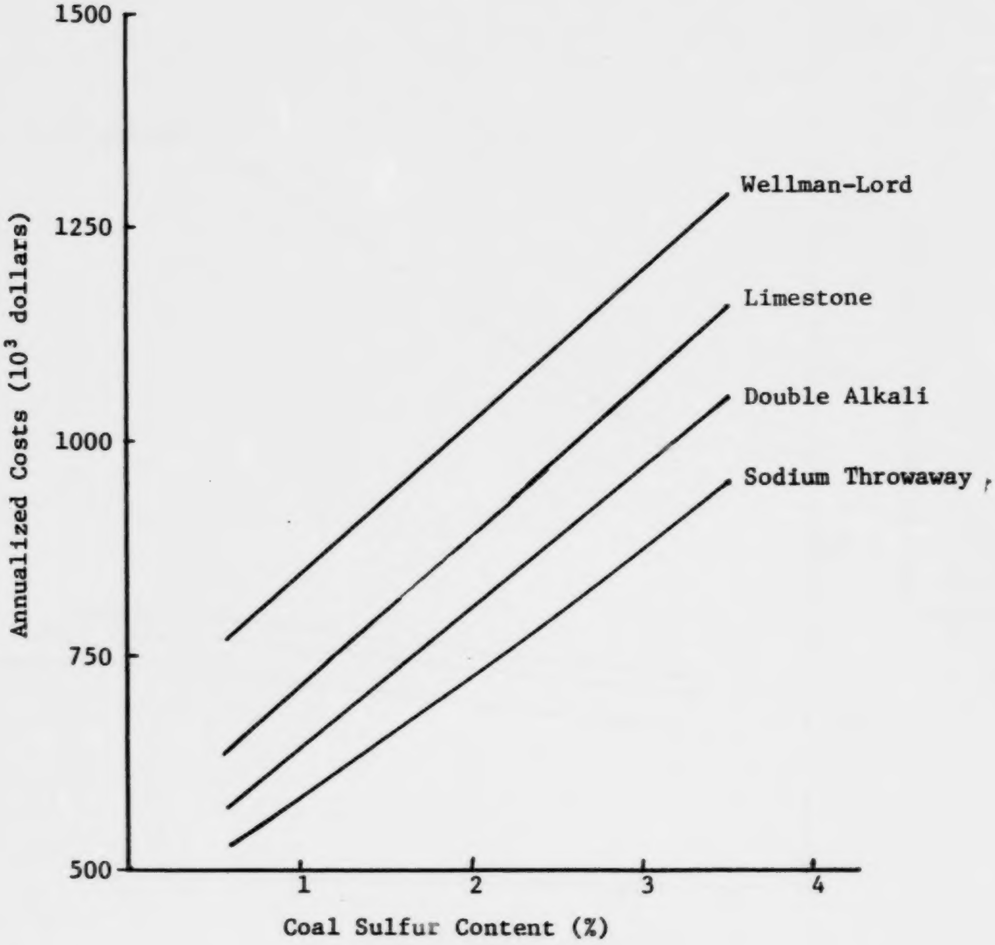

Figure 4.2-6. FGD annualized costs *versus* coal sulfur content. (58.6 MW, 90% removal)

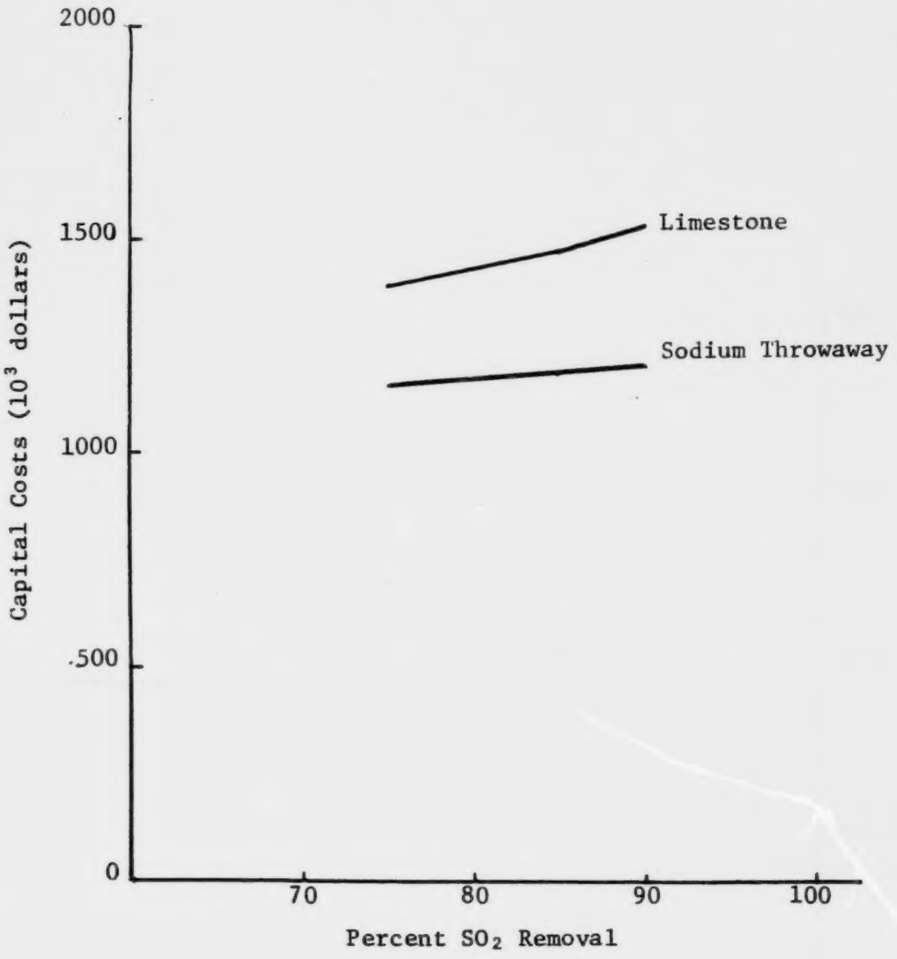

Figure 4.2-7. FGD capital costs *versus* SO_2 removal.
($58.6\ MW_t$ 3.5% S coal)

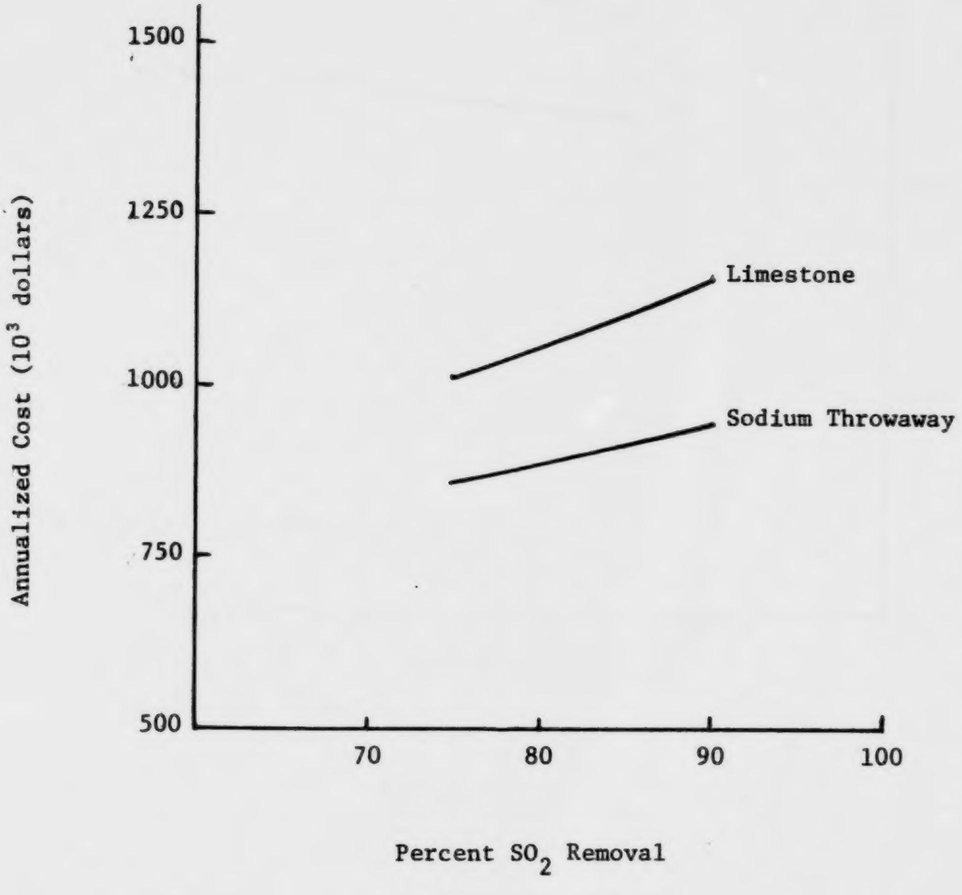

Figure 4.2-8. FGD annualized costs *versus* SO_2 removal. ($58.6\ MW_t$, 3.5% S coal)

Figures 4.2-9 and 4.2-10 illustrate the change in capital and annualized costs with SO_2 removal for the limestone, sodium throwaway, and spray drying processes applied to the 44 MW_t boiler burning 0.6 percent sulfur coal. One can see from these plots the capital intensiveness of the baghouse for the spray drying processes. The limestone and sodium throwaway process capital costs do not include charges for particulate control. With regard to annualized operating costs, however, Figure 4.2-10 illustrates that the relative annualized cost difference between the spray drying process and the sodium and limestone processes is not as great. This is due to the lower disposal costs associated with handling a dry waste product. This figure also illustrates the relative costs of lime and sodium as sorbents for use in the spray drying process.

Cost Effectiveness

The cost effectiveness of the various FGD processes was also determined as part of this study. Cost effectiveness was defined as dollars per kilogram of removed SO_2 (\$/kg SO_2) and was calculated by dividing the annualized process costs by the kilograms of SO_2 removal in a year assuming a 60 percent load factor. Results of these calculations are shown graphically in Figures 4.2-11 through 4.2-15.

Results of these figures show that both coal sulfur content and process size significantly affect the cost effectiveness of an FGD process. For a given size system, cost effectiveness increases with an increasing coal sulfur content. For a fixed coal sulfur content, cost effectivenss increases with increasing process size. Consequently, the most cost effective systems are those designed for large boilers burning high sulfur coal, and the least cost effective systems are those designed for the small boilers burning low sulfur coal. All FGD systems examined here are most cost effective at the stringent level of control, 90 percent SO_2 removal.

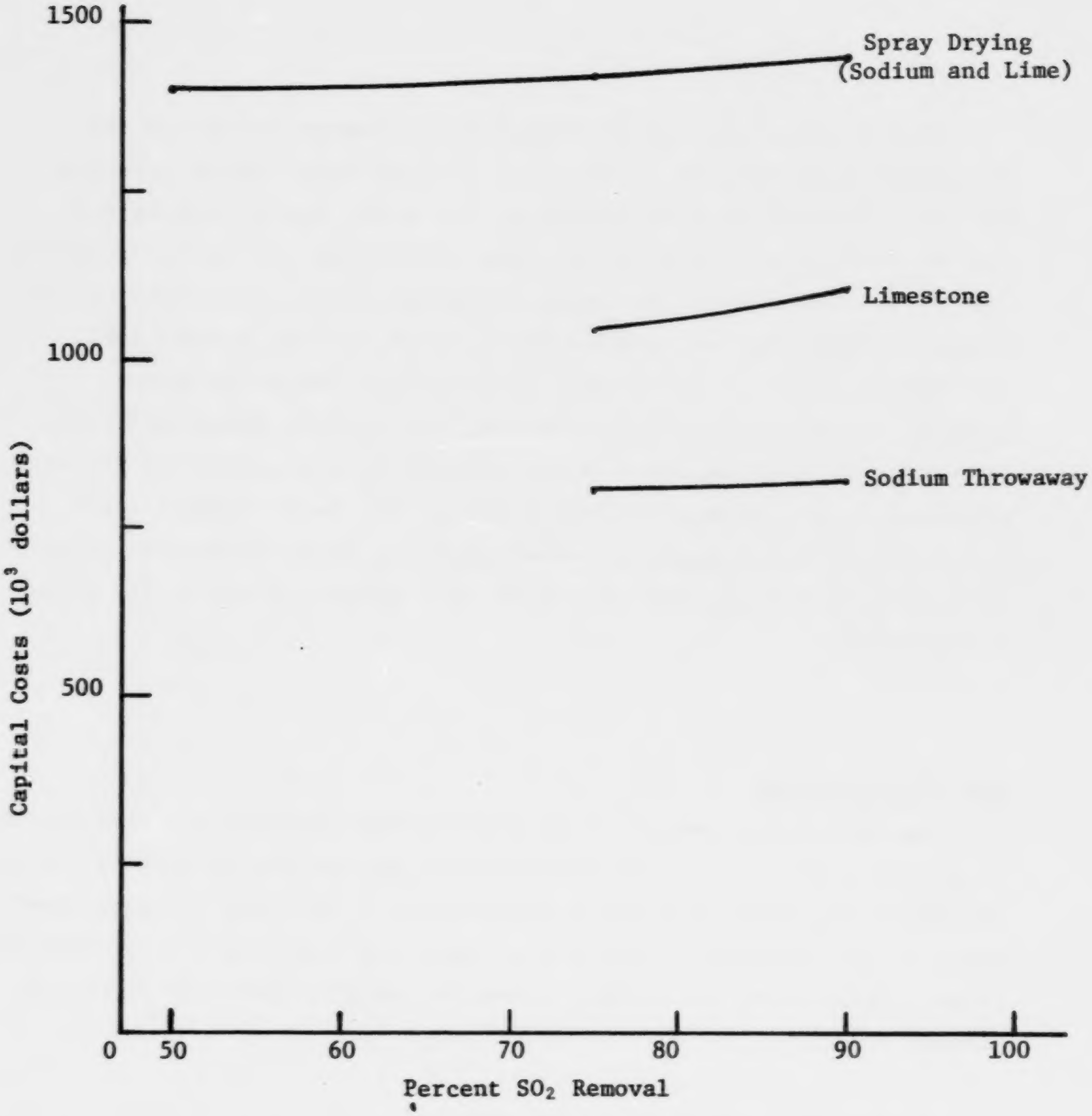

Figure 4.2-9. FGD Capital Costs *versus* SO_2 removal. ($44\ MW_t$, 0.6% S coal)

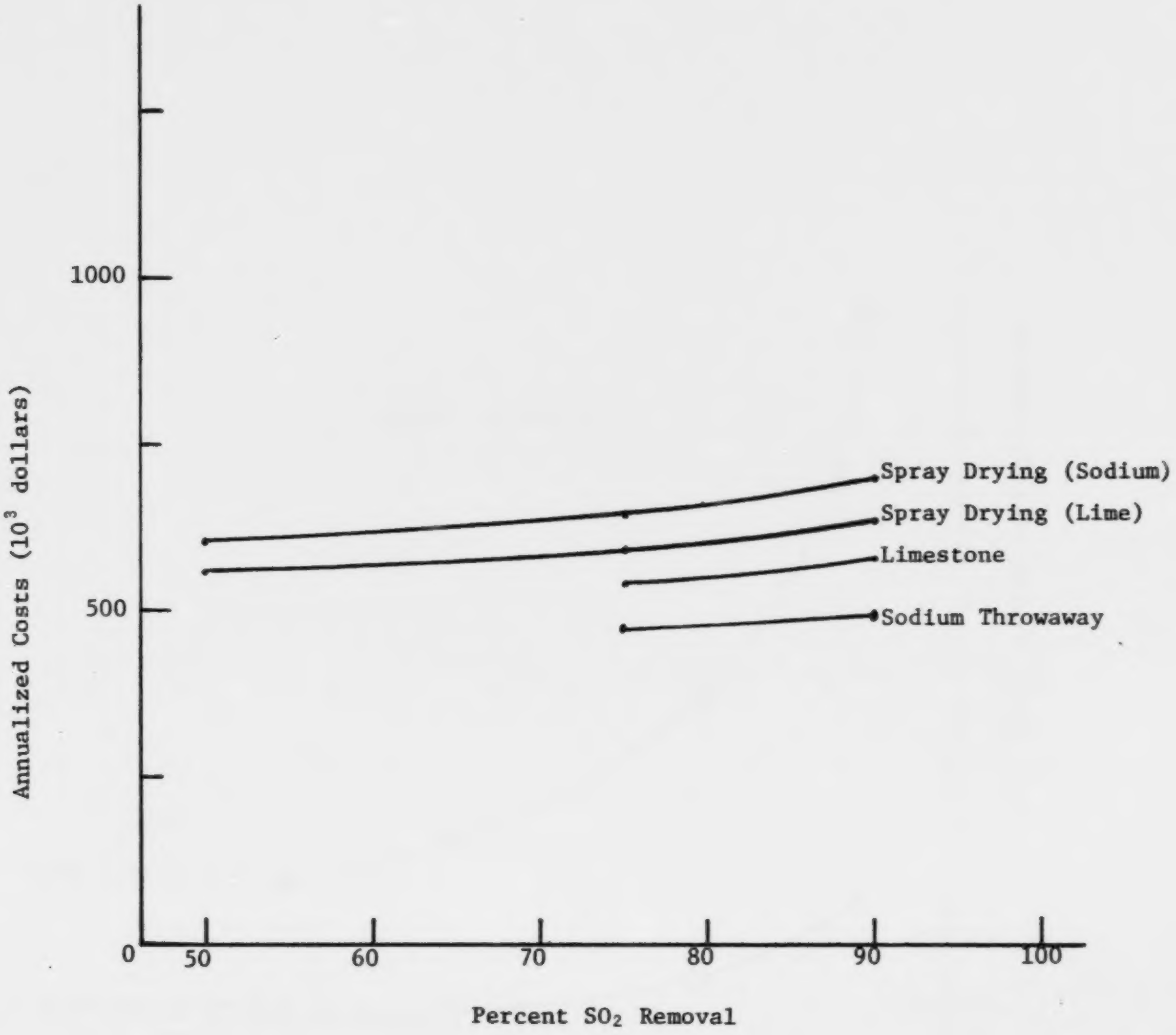

Figure 4.2-10. FGD Annualized Costs *versus* SO_2 removal.
($44\ MW_t$, 0.6% S coal)

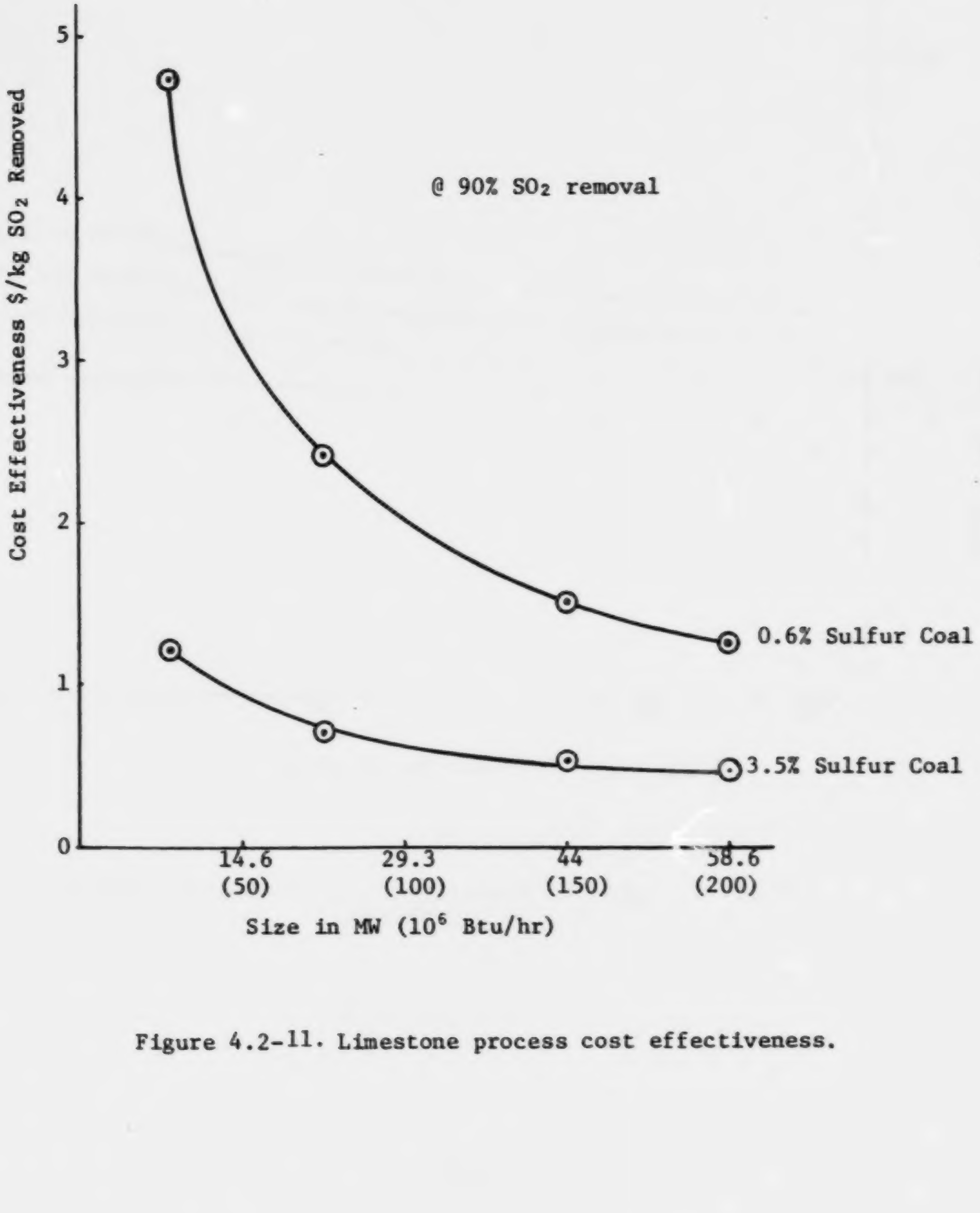

Figure 4.2-11. Limestone process cost effectiveness.

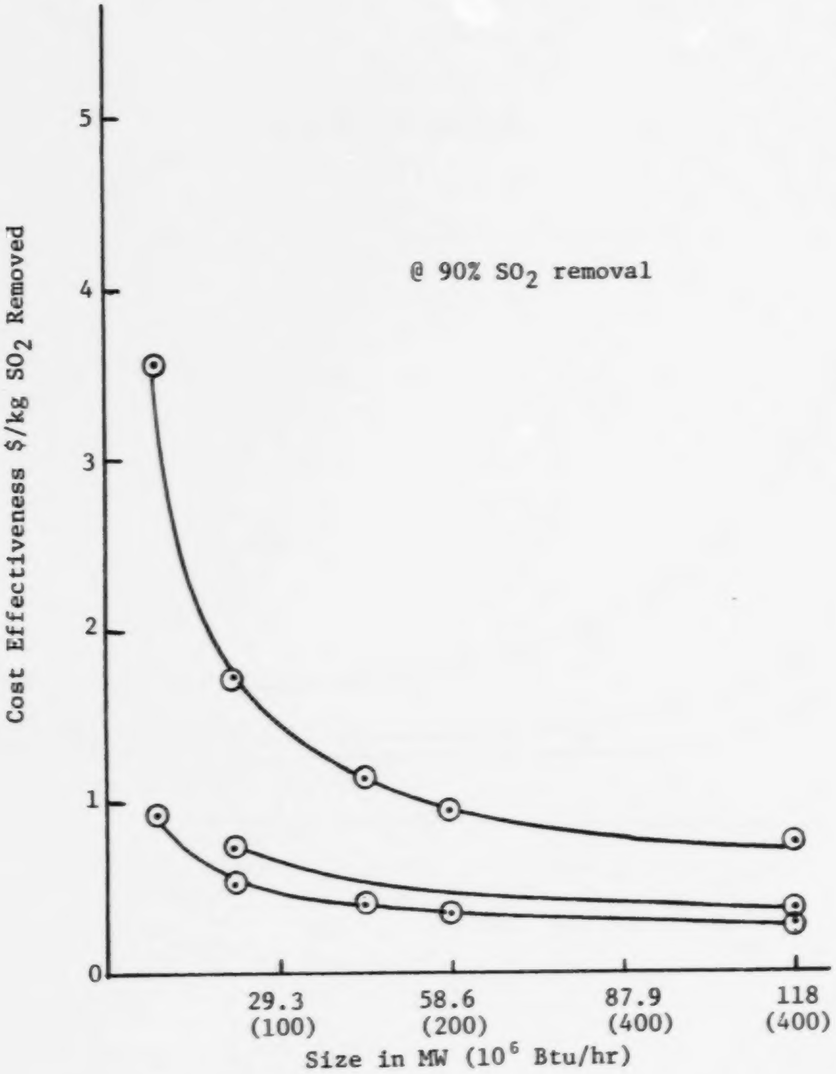

Figure 4.2-12. Sodium throwaway process cost effectiveness.

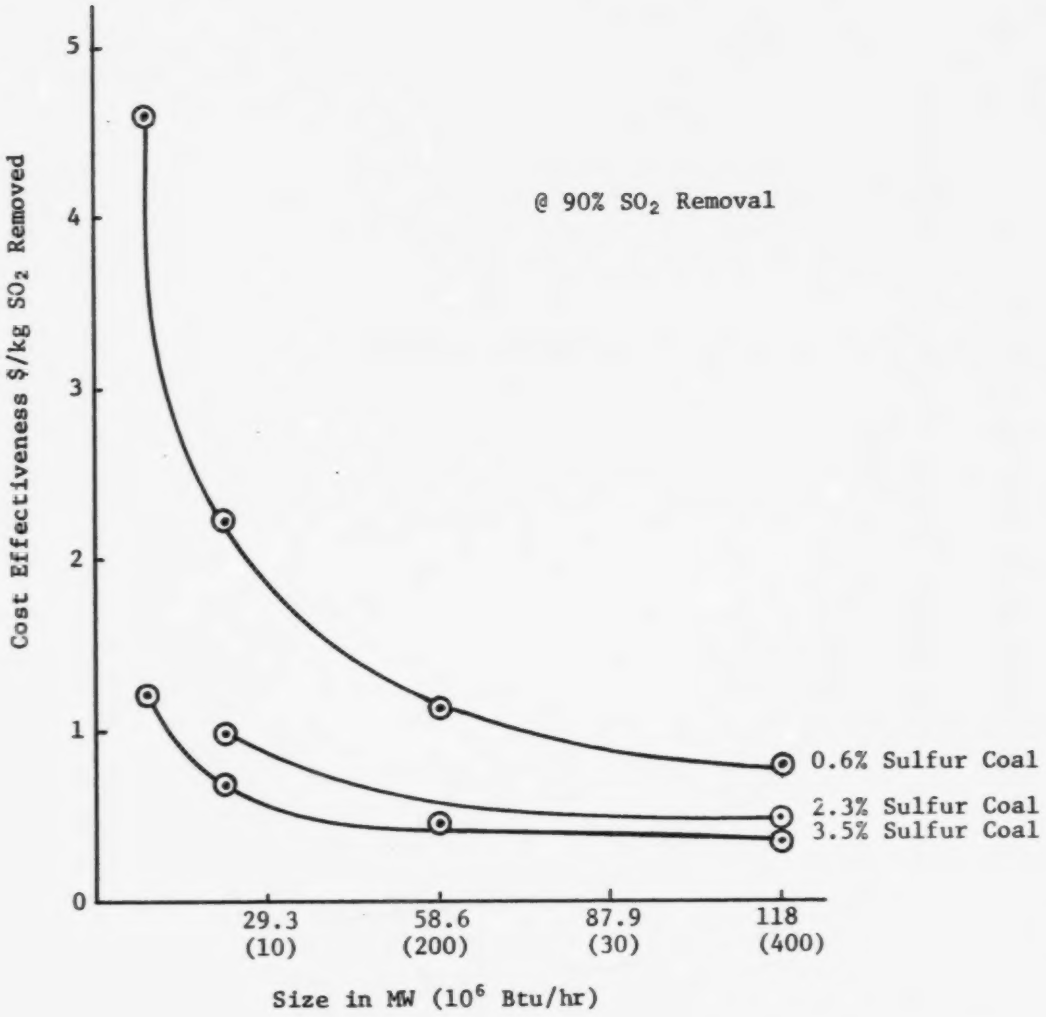

Figure 4.2-13. Double alkali process cost effectiveness.

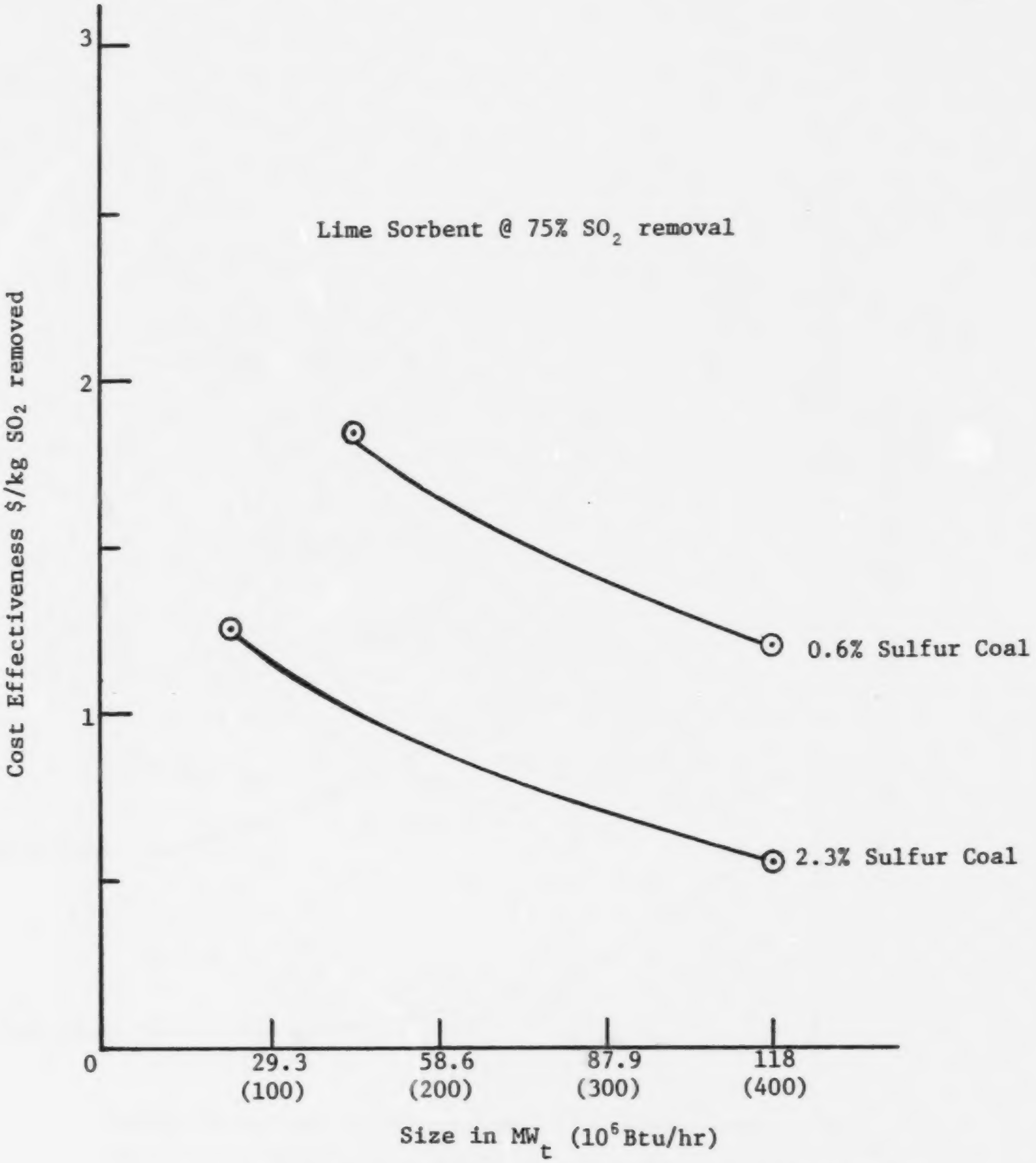

Figure 4.2-14. Spray Drying Process Cost Effectiveness.

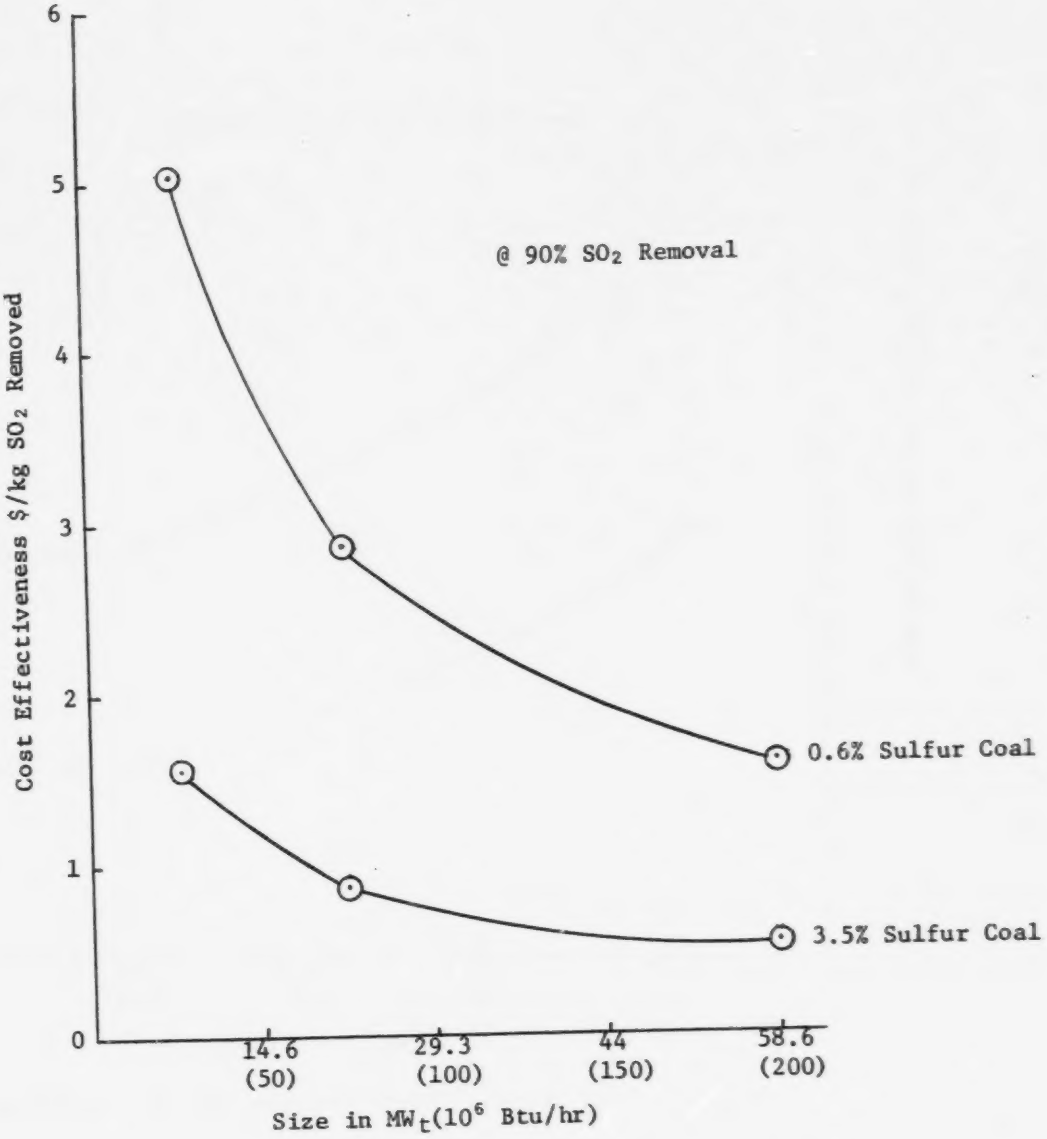

Figure 4.2-15. Wellman-Lord process cost effectiveness.

Retrofit Applications

Application of FGD systems to existing plants usually entails higher costs than those for application to similar new plants. Whereas an FGD system for a new plant can be incorporated into the overall plant design, retrofitting requires that the system be adapted to the given plant configurations; space is often limited, and ongoing plant operations further constrain installation of the system.

Configuration of equipment in the plant sometimes governs the location of the FGD system. For instance, if the boiler stack is on the roof of the boiler house, as it is in many older plants, the FGD system may have to be placed at ground level; this placement could entail long ducting runs from the absorber to the stack or could require construction of a new stack. At some plants the stack is situated directly adjacent to the boiler house or particulate control device, a placement that often necessitates locating the FGD system at some distance. At some plants, especially those located in urban areas, the available space at ground level is inadequate to accommodate the entire FGD system. In such cases either the FGD scrubber units must be stacked, one on top of the other, or additionl land must be acquired adjacent to the plant property.

Terrain of the power plant site also affects the capital cost of the FGD system by sitework and structural requirements. Hilly terrain requires considerable grading and filling to prepare the site for construction of foundations and possible additional structural components. Subsurface conditions can necessitate piling to provide adequate support for the concrete foundations of the FGD system.

Other capital cost components that can be increased because of space restrictions are construction labor and expenses, interest charges during construction (because of longer construction periods), contractor fees and expenses, and allowances for shakedown. Table 4.2-6 summarizes the capital cost impacts of several retrofit requirements.[12]

TABLE 4.2-6. TYPICAL INCREASE IN CAPITAL COSTS WITH VARIOUS RETROFIT REQUIREMENTS[13]

Retrofit requirements	
Long duct runs	4 - 7
Tight space	1 - 18
Delayed construction (1 year delay)	5 - 15
Hilly terrain	0 - 10
New stack	6 - 20
	16 - 70

4.2.2 Example Calculation

The sodium throwaway FGD process was selected for the sample calculation due to its present predominance in industrial boiler installations. The approach used to estimate FGD capital investment costs was as follows. First, a series of material and energy balance calculations were performed to establish process flow rates and energy requirements. Second, each of the FGD processes were divided into a number of process modules which represented separate processing areas. Equipment sizes were then developed for each process module using detailed engineering designs based on the results of the material and energy balances. Finally, capital cost estimates were prepared by contacting process equipment vendors for price quotations in the size range for the standard industrial boilers of this study.

All of the capital costs for each process area are reported in the form of direct capital costs which include all materials and labor installation costs. Particulate control equipment costs were not included in this study except for the spray drying process, and there a baghouse was used for the particulate control device. Table 4.2-7 illustrates the process operating conditions for the example case design, one of the larger FGD systems considered in this study.

TABLE 4.2-7 PROCESS OPERATING CONDITIONS

Operating condition	Value
Boiler combustion rate, MW_t (MBtu/hr)	58.6 (200)
Coal HHV, Btu/lb	11,800
Coal sulfur content, wt%	3.5
Coal ash content, wt%	10.58
Total excess air, %	30
SO_2 feed rate, lb/hr	1126.5
SO_2 recovery, %	90
Flue gas rate; acfm, T°F/psia	74,800; 400/14.7
Year of economics	mid-1978

Raw Material Handling

Raw material handling costs will be a function of the Na_2CO_3 makeup rate. The material balance calculations presented in Appendix A showed a $NaCO_3$ feed rate of 1900 lb/hr for the 58.6 MW_t industrial boiler. The raw material handling costs include costs for the following equipment items:

- pneumatic conveyor (sorbent unloading to storage)
- storage silo (2-week capacity to avert shortages in case of delays in sorbent delivery due to bad weather, strikes, etc.)
- screw feeder (sorbent feed to mixing tank)
- mixing tank (5 minute residence time)
- mixer (agitation for dissolution)
- 2-feed prep pumps (pump feed to circulation tank)

The F.O.B. costs for each of these items were obtained via quotations from process equipment vendors and are presented in Table 4.2-8.

TABLE 4.2-8. RAW MATERIAL HANDLING CAPITAL COSTS

Item	Capacity	Mid-1978 F.O.B. Cost
Pneumatic conveyor	2 +ph (smallest available)	\$ 3,400
Silo	$(1900 \frac{lb}{hr})(24 \frac{hr}{day})(14\ day) = 638{,}000\ lb$	87,000
Screw feeder	2 +ph (smallest available)	1,200
Mixing tank	(123 GPM)(5 min) = 615 gal	1,600
Mixer	615 gal	930
Feed prep pumps (2)	123 GPM ea.	11,000
		\$105,130

$\underline{SO_2\ Scrubbing}$

The F.O.B. costs (acquired from equipment vendors) for the equipment items in the SO scrubbing area are shown below in Table 4.2-9. The capacities are obtained from the material balance presented in Appendix A.

TABLE 4.2-9. SO_2 SCRUBBING CAPITAL COSTS

Item	Capacity	Mid-1978 F.O.B. Cost
Absorber (scrubber)	74,800 acfm	\$72,000
Circulation tank	(748 + 123 GPM)(5 min) = 4360 gal	5,700
Mixer	4360 gal	3,100
Circulation pumps (2)	779 GPM ea.	16,000
		\$96.800

Fans

The F.O.B. costs for the fan (including motor) are shown below in Table 4.2-10.

TABLE 4.2-10 FAN CAPITAL COSTS

Item	Capacity	Mid-1978 F.O.B. Cost
Fan (including motor)	74,800 acfm	$33,000

Wastewater Pumps

The F.O.B. costs for the wastewater pumps (for pumping the wastewater to the boilers' wastewater treating facility) are shown below in Table 4.2-11.

TABLE 4.2-11. WASTEWATER PUMPS CAPITAL COSTS

Item	Capacity	Mid-1978 F.O.B. Cost
Wastewater Pumps (2)	92 GPM ea.	10,000

The F.O.B. costs for each module (process area) are multiplied by an installation factor to obtain the installed costs (including transporation) and are summarized below in Table 4.2-12.

TABLE 4.2-12. INSTALLED COSTS

Process Module	Total F.O.B. Cost	Installation Factor[14]	Installed Cost (10^3\$)
Raw Material Handling	$105,130	1.69	178
SO_2 Scrubbing	96,800	3.11	301
Fans	33,000	2.29	76
Wastewater Pumps	10,400	2.49	26
		Subtotal	581

Costs for utilities and services were estimated at 6% of the installed costs of all process equipment.[15]

$$\text{Utilities and services cost} = .06\ (\$581{,}000)$$
$$= \$35{,}000$$

Installed Cost Summary

For the 58.6 MW_t Case:

Process Area	Costs (10^3 \$)
Raw Material Handling	178
SO Scrubbing	301
Fans	76
Wastewater Pumps	26
Utilities and Services	35
Total	616

All of the indirect capital costs are based upon a percentage of the direct costs as discussed in Section 4.1. Indirect costs include such items as engineering, construction expenses, contractor fees, start-up expenses, and contingencies. Table 4.2-13 illustrates how these indirect capital costs were applied to determine the total turnkey costs (TTC) for the sodium throwaway process.

Land and working capital costs are added to TTC to arrive at Total Capital Investment (TCI). Land is estimated based upon percentages of process costs derived from TVA. Working capital is estimated at 25 percent of total direct operating costs as discussed in Section 4.1 and shown in Table 4.2-13.

Annual process costs for the sodium throwaway process are shown in Table 4.2-14. These costs are based upon a 60 percent stream factor[16] (5256 hrs/yr) @ 100 percent flow. The material balance calculations presented in Appendix A and the energy balance calculations presented in Section 5 provided the basis of the fuel and chemical costs. Prices for labor, raw materials, and utilities were established by PEDCo for consistency[17] and are shown on Table 4.1-2 and on Table 4.2-14.

TABLE 4.2-13

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	178		
SO_2 Scrubbing	301		
Fans	76		
Wastewater Pumps	26		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	39		
Total Direct Costs (TDC)		615	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	62		
Construction Fees (0.1 TDC)	62		
Start-up (0.02 TDC)	12		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		204	
Contingencies (0.2 (TDC+TIC))		164	
TOTAL TURNKEY COSTS (TTC)			983
Land[b] (0.00084 TTC)		0.8	
Working Capital (0.25 Direct Operating Costs)[c]		167	
TOTAL CAPITAL INVESTMENT (TCI)			1151

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

TABLE 4.2-14.
ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			27		
Maintenance Materials (.04 TDC)			27		
Electricity (25.8 mills/kWh)	151	kW	20		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	27.9	m^3/hr	6		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	20.9	m^3/hr	11		
Solids Disposal ($.044/kg)	-	kg/hr	-		
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	864	kg/hr	450		
Total Direct Operating Cost				667	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			47		
Total Overhead Costs				85	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				196	
TOTAL ANNUALIZED COSTS					947
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

Overhead costs were estimated as percentages of direct labor and maintenance costs as discussed in Section 4.1.[18] Both payroll and plant overhead expenses were estimated as shown in Table 4.2-14.

Capital charges of 17 percent of the Total Capital Investment TCI are based upon a 10 percent return on investment, 15 yr plant life, straight-line depreciation, and 4 percent G&A expenses. The sum of the direct operating costs, overhead costs, and capital charges is the total annualized cost. In the cases presented, all costs, both capital and annual, are expressed in thousands of dollars. Cost tables for the other cases considered are found in Appendix B.

4.3 COSTS TO CONTROL OIL-FIRED BOILERS

Both capital and annualized costs of controlling SO_2 emissions from the standard residual oil-fired boiler were estimated using the same cost estimating procedure as described in Section 4.1. Costs were developed for the limestone process in order to determine the impact of treating flue gas from a residual oil-fired boiler. For comparison, costs were also developed for a coal-fired system of the same size treating flue gas from a 3.5 percent sulfur coal. Table 4.3-1 summarizes the results of these calculations.

TABLE 4.3-1. COST COMPARISON OF OIL- AND COAL-FIRED LIMESTONE FGD COSTS

Boiler size	Fuel type	Percent SO_2 removal	Capital investment costs (10^3 \$)	Total annualized costs (10^3 \$)
44 MW_t (150×10^6 Btu/hr)	Eastern Coal 3.5% S	90	1385	974
		75	1270	865
	Resid Oil 3.0% S	90	1017	742
		75	942	683

The major differences between the oil- and coal-fired systems are that the oil-fired system processes less gas and removes a small quantity of SO_2 than the coal-fired system. This is because the oil-fired boiler operates with 15 percent excess air versus 50 percent for the standard spreader stoker boiler, and the uncontrolled SO_2 emissions from the oil-fired boiler are 471 kg/hr versus 846 kg/hr for the coal-fired system. These differences result in the oil-fired system having smaller raw material feed equipment, smaller gas handling equipment, and smaller solids handling equipment.

Figure 4.3-1 illustrates the difference in the oil- and coal-fired costs graphically. Costs of the coal-fired system are about one-third higher for a given SO_2 removal. Similar cost differentials should exist for the other processes.

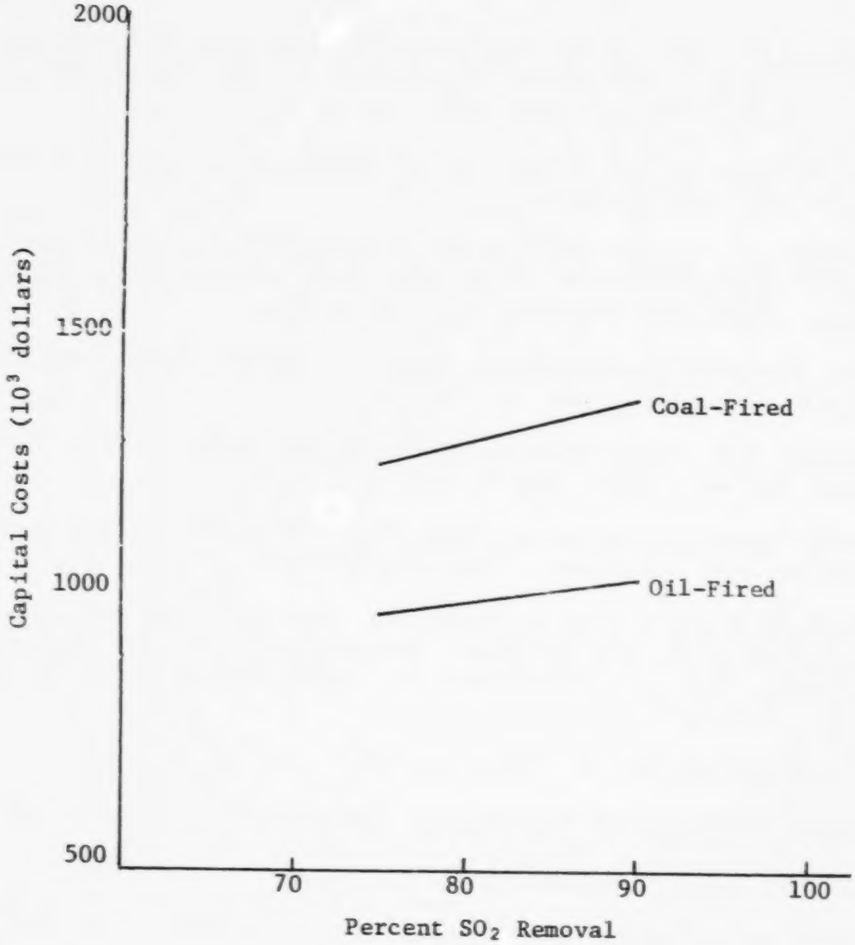

Figure 4.3-1. Comparison of oil- and coal-fired FGD costs. (Limestone Process, 44 MW_t)

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SECTION 5

ENERGY IMPACT OF CANDIDATES FOR BEST EMISSION CONTROL SYSTEMS

5.1 INTRODUCTION

This section addresses the energy requirements associated with the control of SO_2 emissions from small industrial boilers. Five FGD systems were selected as being the best candidate SO_2 control systems and the energy requirements associated with each are shown in Table 5.1-1. These energy requirements compare quite well with FGD energy requirements presented in a recent study prepared by Rubin[1] and with energy requirements reported by FGD system operators at a recent symposium on FGD energy requirements held at Lehigh University.[2] Both of these sources have reported energy requirements of limestone systems with stack gas reheat to vary from about 3 to 3.5 percent of the net heat input to the boiler.

TABLE 5.1-1. RANGE OF FGD SYSTEM ENERGY REQUIREMENTS*

SO_2 control method	Energy Requirement Not Including Reheat	Energy Requirement Including Reheat
Limestone	0.9-1.8	2.4-3.5
Sodium Scrubbing	0.5-0.8	2.0-2.6
Double Alkali	0.5-0.6	2.0-2.3
Spray Drying	0.5-0.8	0.5-0.8
Wellman-Lord	1.6-6.2	3.2-7.9

* Energy Requirements expressed as percent of net heat input to boiler.

The variations in energy requirements for these processes are due to different levels of sulfur in the coal, different levels of SO_2 control, and to a smaller extent, unit size.

There are four major energy consumption areas that are used for all of these processes:

- Raw materials handling and fuel preparation
- SO_2 scrubbing
- Fans
- Utilities and services

A fifth area of energy consumption, stack gas reheat, was not considered as part of the industrial boiler FGD systems' energy requirements since the majority of industrial boiler FGD applications do not reheat the exhaust gas. To illustrate the impact of reheat, Table 5.1-1 was prepared assuming reheat of the stack gases to 175°F. This table shows that stack gas reheat creates an energy penalty of 1.5-1.7 percent of the net heat input to the boiler and significantly increases the energy consumption of all wet scrubbing processes. The spray drying process exhausts its stack gas at 175°F and, consequently, no reheat would be required.

In addition to the above energy consumption areas, the sodium scrubbing process has energy requirements for wastewater disposal and the Wellman-Lord process requires low pressure steam as part of its regeneration operations and methane for sulfur production. No energy impact for disposal was charged to the three other throwaway processes as the waste is a solid and was assumed to be hauled off-site and disposed of by a contractor. Each of the above process areas will be discussed briefly below as they relate to FGD energy requirements.

The raw material handling and feed preparation operation is designed to receive, store and prepare makeup reagents for the FGD processes. This requires storage silos, conveyors, mixers, slurry or solution preparation tanks and pumps. The double-alkali process uses lime as the regenerant or precipitation reagent. Facilities for calcining limestone to produce the lime were not included in the design basis. If these facilities were included, the energy requirements of the raw material handling and feed preparation operations would increase and would impact the total energy requirements of the double-alkali process with an approximate 25 percent increase.[3]

The major energy requirement associated with the SO_2 scrubbing process area is electricity to run the flue gas fans and process recirculation pumps. Liquid circulation (L/G) rates for these energy consumption calculations were based on information provided in Section 2 of this report for each of the candidate processes. The pressure drop to be overcome for each case was estimated by using the following empirical relationship developed by TVA.[4]

$$\Delta P = 1.68 + 7.17 \times 10^{-5} \text{ (Bed Height)} \times L/A \times V^2 \qquad (5\text{-}1)$$

where:

ΔP = Pressure drop (in H_2O)
Bed Height = Total height of TCA spheres (in)
L = Liquid circulation rate (gpm)
A = Scrubber area (ft^2)
V = Gas velocity (ft/sec)

To evaluate the energy requirement for installations using reheat, stack gas reheat was assumed to occur in indirect steam reheaters. The minimum stack exit temperature required to prevent sulfuric acid mist formation and provide plume buoyancy is not well defined. Common practice, and the design basis for these additional calculations is to reheat the scrubbed gas to achieve an exit temperature of 353°K (175°F) which requires ∿50°F of reheat.[5]

Utilities and services such as instrument air, lighting, heating, cooling, etc. are also required for each FGD process. For this analysis the amount of energy consumed for utilities and services was estimated from process energy requirements developed for utility boilers.

The limestone, double-alkali, sodium scrubbing, and spray drying FGD processes produce waste materials for disposal. Ponding is the normal on-site method of FGD waste disposal. The processes with a solid phase waste material were assumed to use a contractor for waste removal and disposal so no energy penalty was charged. It was assumed that the liquid wastes from the sodium scrubbing process would be pumped one mile to a central on-site wastewater treatment plant.

The Wellman-Lord process produces a concentrated SO_2 stream from regeneration of the SO_2 scrubbing liquor. A set of evaporator/crystallizers are used to regenerate the SO_2 scrubbing liquor. This equipment uses low pressure steam as the energy source. Because the regeneration operation produces a concentrated SO_2 stream, conversion to elemental sulfur or sulfuric acid is possible. For this study, a proprietary process of the Allied Chemical Company which uses methane as a reductant was selected as the basis for converting the SO_2 to elemental sulfur.[6]

5.2 ENERGY IMPACT OF CONTROLS FOR COAL-FIRED BOILERS

A series of material balance calculations were performed for each of the FGD systems to identify stream flow rates, equipment sizes, effluent streams, and raw material requirements. Results of these calculations are presented in Appendix A. Process energy requirements were then calculated using the material balance results and the assumptions and design bases listed in Table 5.2-1. Process energy requirements calculated for each of the FGD systems under consideration are shown in Tables 5.2-2 through 5.2-6.

Results of these calculations indicate that the process energy penalties range from about 0.5 to 2 percent of the gross heat input to the boiler for the three throwaway processes and from about 2 to 6 percent for the Wellman-Lord process. The larger energy consumption for the Wellman-Lord process is due to the steam and methane requirements for the regeneration and SO_2 reduction portions of the process. Energy penalties associated with the SO_2 reduction section of the Wellman-Lord process were estimated from the results of a previous study.[7]

TABLE 5.2-1. MATERIAL AND ENERGY BALANCE ASSUMPTIONS AND DESIGN BASES

Process Parameters	FGD PROCESSES					
	Limestone	Sodium Throwaway	Wellman-Lord	Double Alkali	Spray Dryer Sodium	Spray Dryer Lime
L/G ℓ/m^3, (gal/10^3 acf)[1]	A (A)	1335 (10)	1335 (10)	1335 (10)	40 (0.3)	40 (0.3)
Particulate Removal[2]	99 Percent Upstream of FGD System				Baghouse Downstream of Spray Dryer	
Stoichiometry (moles sorbent/mole sorbed SO_2)[1]	1.2	B	1.0	1.0	C	C
Gas Pressure Drop Pa (in H_2O)[3]	A (A)	100 (8)	100 (8)	100 (8)	75 (6)	75 (6)
Pump Discharge Pressure Pa (psi)[2]	5227 (15)	3484 (10)	3484 (10)	3484 (10)	3484 (10)	3484 (10)
Pumping Height M (ft)[2]	6 (20)	6 (20)	6 (20)	6 (20)	6 (20)	6 (20)

(A) Limestone System L/G and ΔP

Coal type	Percent SO_2 removal 90	85	75
Eastern (3.5% S)			
L/G	10.7 (80)	8.0 (60)	5.3 (40)
ΔP	214 (17)	176 (14)	151 (12)
Western (0.6% S)			
L/G	9.3 (70)	7.3 (55)	5.3 (40)
ΔP	189 (15)	164 (13)	151 (12)

(B) Sodium Throwaway Stoichiometry

90% Removal - 1.05
85% Removal - 0.95
75% Removal - 0.85

(C) Spray Dryer Stoichiometry

	Sodium[4]	Lime[5]
90% Removal -	1.1	2.0
75% Removal -	0.8	1.2
50% Removal -	0.5	0.65

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[1] Based on process data presented in Section 2.
[2] Radian assumption.
[3] Based on TVA empirical relationship.
[4] Reference 8.
[5] Reference 9.

TABLE 5.2-2. ENERGY REQUIREMENTS FOR THE LIMESTONE FGD PROCESS

			Energy consumed in control device (kW_t)						
Boiler size and type	Coal type	SO_2 removal level	Raw material handling and preparation	Liquid	Fan	Utilities & services	Total kW_t	Total 10^6 Btu/hr	Percent increase in energy use over uncontrolled boiler
8.8 MW_t ($30x10^6$ Btu/hr) Underfeed stoker	Eastern 3.5% sulfur	90	17.0	54.8	81.3	1.5	154.6	0.53	1.8
		85	16.1	41.1	67.0	1.5	125.7	0.43	1.4
		75	14.0	27.8	57.3	1.5	100.6	0.34	1.1
22 MW_t ($75x10^6$ Btu/hr) Chaingrate stoker	Eastern 3.5% sulfur	90	42.7	136.7	203.7	3.7	386.8	1.32	1.8
		75	35.7	69.7	143.8	3.7	252.9	.86	1.2
44 MW_t ($150x10^6$ Btu/hr) Spreader stocker	Eastern 3.5% sulfur	90	86.1	274.6	408.7	7.4	776.8	2.65	1.8
		75	71.4	140.0	288.5	7.4	507.3	1.73	1.2
58.6 MW_t ($200x10^6$ Btu/hr) Publverized coal	Eastern 3.5% sulfur	90	114.2	317.4	471.7	8.6	911.9	3.11	1.6
		85	108.3	240.0	388.3	8.6	745.2	2.54	1.3
		75	95.4	162.2	333.0	8.6	599.2	2.05	1.0
8.8 MW_t ($30x10^6$ Btu/hr) Underfeed stoker	Western .6% sulfur	90	3.5	45.7	72.7	1.5	123.4	0.42	1.4
		85	3.4	36.0	63.0	1.5	103.9	0.35	1.2
		75	3.0	27.8	58.2	1.5	90.5	0.31	1.0
22 MW_t ($75x10^6$ Btu/hr) Underfeed stoker	Western .6% sulfur	90	9.0	115.4	183.0	3.8	311.2	1.06	1.4
		75	7.6	66.5	146.5	3.8	224.4	0.77	1.0
44 MW_t ($150x10^6$ Btu/hr) Spreader stoker	Western .6% sulfur	90	18.2	230.1	364.8	7.6	620.7	2.12	1.4
		75	15.2	132.9	292.0	7.6	447.7	1.53	1.0
58.6 MW_t ($200x10^6$ Btu/hr Pulverized coal	Western .6% sulfur	90	23.9	267.6	423.7	8.8	724.0	2.47	1.2
		85	22.7	210.8	367.2	8.8	609.5	2.08	1.0
		75	20.2	154.0	339.0	8.8	522.0	1.78	0.9

TABLE 5.2-3. ENERGY REQUIREMENTS FOR THE SODIUM THROWAWAY FGD PROCESS

Boiler size and type	Coal type	SO_2 removal level	Energy consumed in control device (kW_t)							Percent increase in energy use over uncontrolled boiler
			Raw material handling and preparation	Liquid	Fan	Disposal	Utilities & services	Total kW_t	Total 10^6 Btu/hr	
8.8 MW_t (30x10^6 Btu/hr) Underfeed stoker	Eastern 3.5 sulfur	90	1.4	6.7	38.2	23.3	1.6	71.2	0.24	0.8
		85	1.3	6.7	38.2	21.7	1.6	69.5	0.24	0.8
		75	1.2	6.7	38.2	19.5	1.6	67.2	0.23	0.8
22 MW_t (75x10^6 Btu/hr) Chaingrate stoker	Eastern 3.5% sulfur	90	3.5	17.3	95.8	60.3	4.0	180.9	0.62	0.8
		75	2.9	16.7	95.8	49.0	4.0	168.4	0.57	0.8
44 MW_t (150x10^6 Btu/hr) Spreader stoker	Eastern 3.5% sulfur	90	7.0	34.6	192.3	121.0	8.0	362.9	1.24	0.8
		75	5.8	33.4	192.3	97.4	8.0	336.9	1.15	0.8
58.6 MW_t (200x10^6 Btu/hr) Pulverized coal	Eastern 3.5% sulfur	90	9.3	40.7	222.0	160.4	9.3	441.7	1.51	0.8
		85	8.8	40.1	222.0	145.4	9.3	425.6	1.45	0.7
		75	7.7	39.5	222.0	130.2	9.3	408.7	1.39	0.7
118 MW_t (400x10^6 Btu/hr) Pulverized coal	Eastern 3.5% sulfur	90	18.6	80.9	433.0	322.7	17.7	872.9	2.98	0.7
22 MW_t (75x10^6 Btu/hr) Chaingrate stoker	Medium sulfur 2.3% sulfur	90	2.0	16.2	95.7	34.9	3.9	152.7	0.52	0.7
118 MW_t (400x10^6 Btu/hr) Pulverized coal	Medium sulfur 2.3% sulfur	90	10.9	75.3	438.1	188.4	17.9	730.6	2.49	0.6
8.8 MW_t (30x10^6 Btu/hr) Underfeed stoker	Western .6% sulfur	90	0.3	5.8	38.7	5.1	1.6	51.5	0.18	0.6
		85	0.3	5.8	38.7	4.5	1.6	50.9	0.17	0.6
		75	0.2	5.8	38.7	4.2	1.6	50.5	0.17	0.6
22 MW_t (75x10^6 Btu/hr) Chaingrate stoker	Western .6% sulfur	90	0.7	14.3	97.7	12.7	4.1	129.5	0.44	0.6
		75	0.6	14.3	97.7	10.3	4.1	127.0	0.43	0.6
44 MW_t (150x10^6 Btu/hr) Spreader stoker	Western .6% sulfur	90	1.5	28.7	194.7	25.4	8.2	258.5	0.88	0.6
		75	1.2	28.7	194.7	20.6	8.2	253.4	0.86	0.6
58.6 MW_t (200x10^6 Btu/hr) Pulverized coal	Western .6% sulfur	90	1.9	34.0	226.0	34.0	9.5	305.4	1.04	0.5
		85	1.9	33.7	226.0	30.7	9.5	301.8	1.03	0.5
		75	1.7	33.7	226.0	27.4	9.5	298.1	1.02	0.5
118 MW_t (400x10^6 Btu/hr) Pulverized coal	Western .6% sulfur	90	3.9	67.2	445.6	[illegible]	18.4	603.1	2.06	0.5

TABLE 5.2-4. ENERGY REQUIREMENTS FOR THE DOUBLE-ALKALI PROCESS

			Energy consumed in control device (kW_t)						
		SO_2 removal level	Raw material handling and preparation			Utilities & services	Total		Percent increase in every use over uncontrolled boiler
Boiler size and type	Coal type			Liquid	Fan		kW	10^6 Btu/hr	
8.8 MW_t (30x10^6 Btu/hr) Underfeed stoker	Eastern 3.5% sulfur	90	1.9	10.4	38.2	1.6	52.1	0.18	0.6
	Western .6% sulfur	90	0.4	6.2	38.7	1.6	46.9	0.16	0.5
22 MW_t (75x10^6 Btu/hr) Chaingrate stoker	Eastern 3.5% sulfur	90	4.8	26.1	94.3	3.8	129.5	0.44	0.6
	Medium sulfur 2.3% sulfur	90	2.7	20.8	95.7	3.9	123.1	0.42	0.6
	Western .6% sulfur	90	0.9	15.5	97.7	4.1	118.2	0.40	0.5
58.6 MW_t (200x10^6 Btu/hr) Pulverized coal	Eastern 3.5% sulfur	90	12.7	63.2	222.0	9.3	307.2	1.05	0.5
	Western .6% sulfur	90	2.4	36.9	226.0	9.5	274.8	0.94	0.5
118 MW_t (400x10^6 Btu/hr) Pulverized coal	Eastern 3.5% sulfur	90	25.5	129.4	433.0	17.7	605.6	2.07	0.5
	Medium sulfur 2.3% sulfur	90	14.2	100.6	438.1	17.9	570.8	1.95	0.5
	Western .6% sulfur	90	4.8	73.3	446.6	18.4	543.1	1.85	0.5

TABLE 5.2-5. SPRAY DRYING ENERGY REQUIREMENTS

Boiler size		Coal sulfur content	SO_2 removal level (%)	Type alkali used	Energy consumed in control device, kW_t					Total		Percent increase in energy use over uncontrolled boiler
MW_t	(10^6 Btu/hr)				Raw materials handling and preparation	Liquid pumping	Fan	Atomization	Utilities and services	kW_t	10^6 Btu/hr	
17.5	(60)	0.6	75	Sodium	0.6	0.2	55.6	71.8	3.2	131.4	0.45	0.7
44	(150)	0.6	90	Sodium	1.4	0.5	139.7	122.3	8.2	272.1	0.93	0.6
44	(150)	0.6	75	Sodium	1.2	0.5	139.7	121.3	8.2	270.9	0.92	0.6
44	(150)	0.6	50	Sodium	0.8	0.5	139.7	120.2	8.2	269.4	0.92	0.6
58.6	(200)	0.6	75	Sodium	1.6	0.6	162.8	132.9	9.7	307.6	1.05	0.5
22	(75)	2.3	70	Lime	2.0	0.4	70.7	106.0	3.9	183.0	0.62	0.8
44	(150)	0.6	90	Lime	1.8	0.5	139.7	121.3	8.2	271.5	0.93	0.6
44	(150)	0.6	75	Lime	1.5	0.5	139.7	120.0	8.2	269.9	0.92	0.6
44	(150)	0.6	50	Lime	1.0	0.5	139.7	119.1	8.2	268.5	0.92	0.6
118	(400)	0.6	70	Lime	3.8	1.8	330.5	246.6	18.4	601.1	2.05	0.5
118	(400)	2.3	70	Lime	10.6	1.9	323.7	257.2	17.9	611.3	2.09	0.5

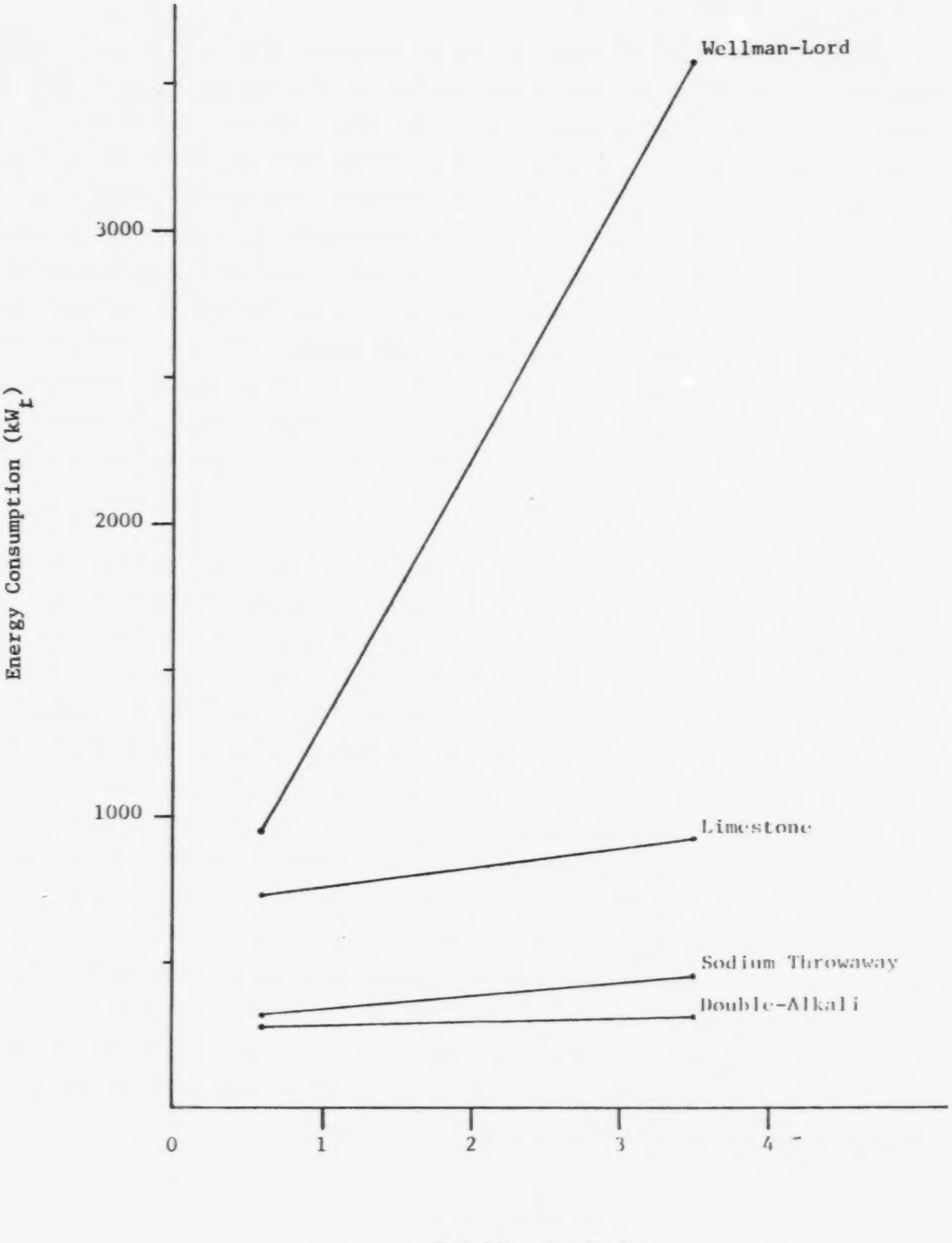
TABLE 5.2-6. ENERGY REQUIREMENTS FOR THE WELLMAN-LORD PROCESS

Boiler size and type	Coal type	SO_2 removal level	Energy consumed in control device (kW_t)								Percent increase in energy use over uncontrolled boiler
			Raw materials handling and preparation	Liquid[a]	Fan	Process steam	Methane	Utilities & services	Total		
									kW	10^6 Btu/hr	
8.8 MW_t (30×10^6 Btu/hr) Underfeed stoker	Eastern 3.5% sulfur	90	1.4	11.5	35.2	337	155	1.7	541.3	1.85	6.2
	Western .6% sulfur	90	.3	6.8	35.7	70	35	1.7	149.5	0.51	1.7
22 MW_t (75×10^6 Btu/hr) Chaingrate stoker	Eastern 3.5% sulfur	90	3.5	31.2	88.5	844	392	4.3	1363.5	4.65	6.2
	Western .6% sulfur	90	.7	16.3	90.2	179	82	4.4	372.6	1.27	1.7
58.6 MW_t (200×10^6 Btu/hr) Pulverized coal	Eastern 3.5% sulfur	90	9.3	82.9	205.5	2220	1048	9.6	3575.3	12.20	6.1
	Western .6% sulfur	90	1.9	42.2	208.6	469	219	10.2	950.9	3.25	1.6

[a] Includes electricity for operation of SO_2 scrubbing pumps and pumps for the SO_2 reduction process area.

Figure 5.2-1 which is based on the calculations developed for this study illustrates the effect of coal sulfur content on process energy requirements. This figure shows a larger energy requirement for a limestone system applied to a 3.5 percent sulfur coal than for a system applied to an 0.6 percent sulfur coal. The energy increase, about 12 percent, is primarily due to the increased energy requirements of the feed preparation area since more alkali is required for the higher sulfur coal. In addition, higher liquid pumping rates and increased system pressure drop for the 3.5 percent sulfur coal case also contribute to the increased energy requirements. The double alkali and sodium throwaway systems show even less of a variation in energy consumption with coal sulfur content. The most noticeable change is with the Wellman-Lord system which is due primarily to increases in the regeneration and SO_2 reduction sections of the process.

Figures 5.2-2 and 5.2-3 illustrate the effect of boiler size on process energy requirements. In all cases, the amount of energy consumed by an FGD process increases with increasing boiler capacity. If the FGD energy consumption is plotted as a percent of the heat input to the boiler, however, a different result is shown. Figures 5.2-4 and 5.2-5 illustrate that except for the large sized boilers, the boiler size has essentially no effect on the energy penalties when they are expressed as a percent of boiler heat input. In this figure, the percent energy penalty for each of these processes is shown to decrease in going to the large 58.6 MW (200 X 10^6 Btu/hr) size. The reason for the decrease is the relatively lower fan energy requirements for the larger sized standard boilers which were used for the basis of this study because the percent excess air decreases from 50 percent down to 30 percent as specified by the standard boiler characterizations.[10] Figures 5.2-3 and 5.2-5 also illustrate that the energy requirements for a spray drying process are essentially the same as those for a sodium throwaway process when reheat is required.



% Sulfur in Coal
(58.6 MW$_t$ Boiler Size; 905 SO_2 Removal)

Figure 5.2-1. Energy Consumption versus coal sulfur content.

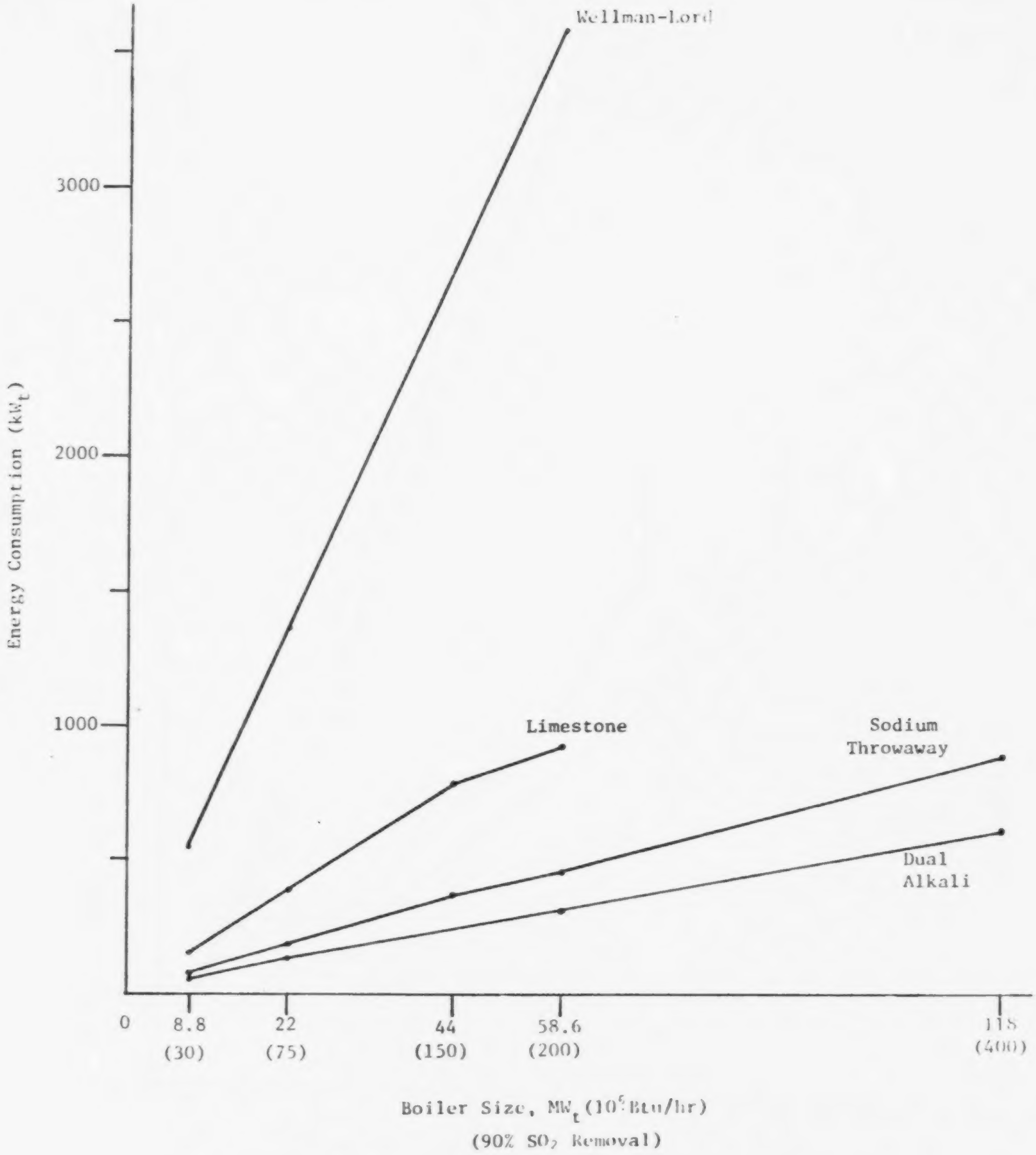


Figure 5.2-2. Energy consumption versus boiler size: high sulfur eastern coal

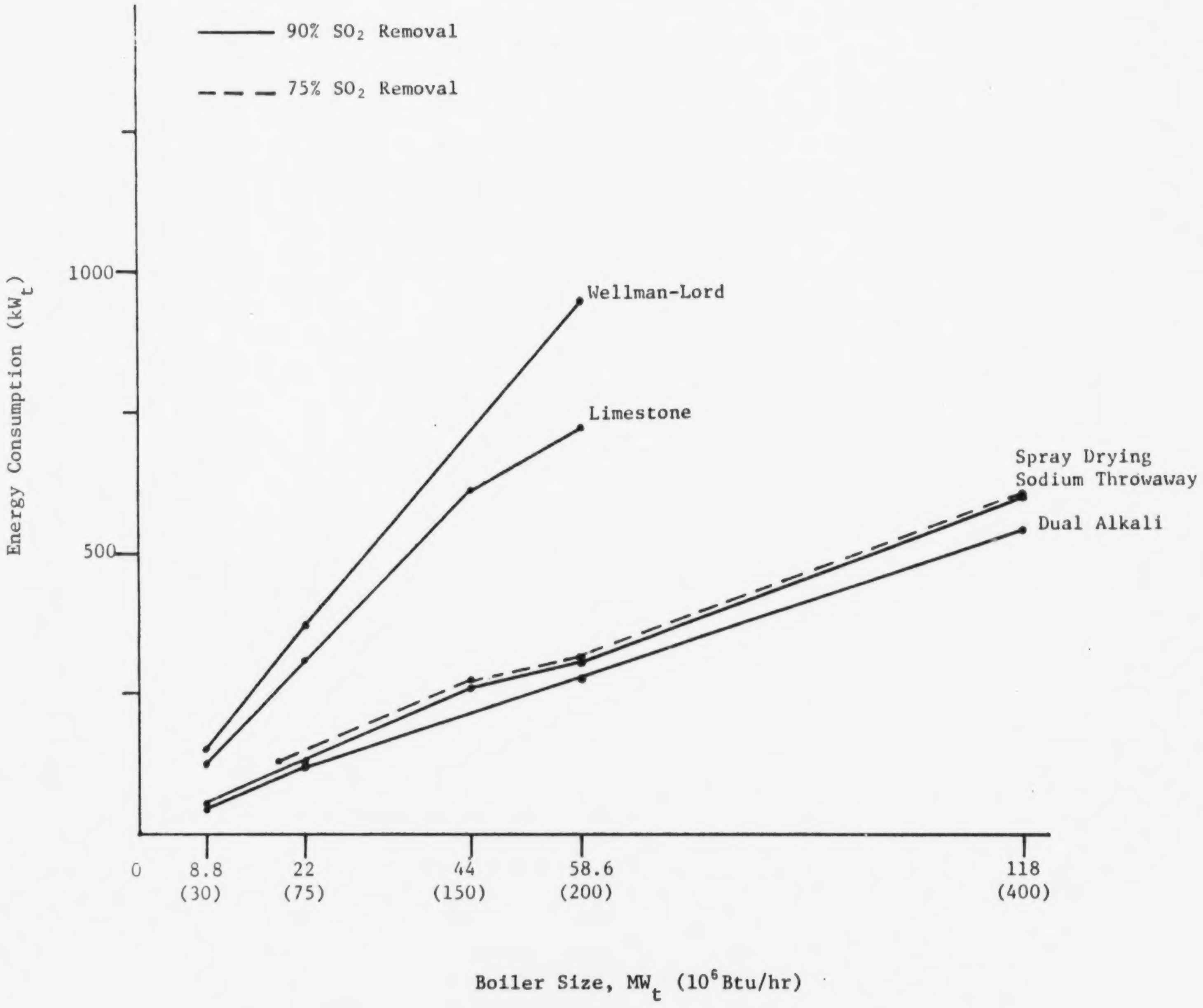


Figure 5.2-3. Energy Consumption versus boiler size: low sulfur western coal.

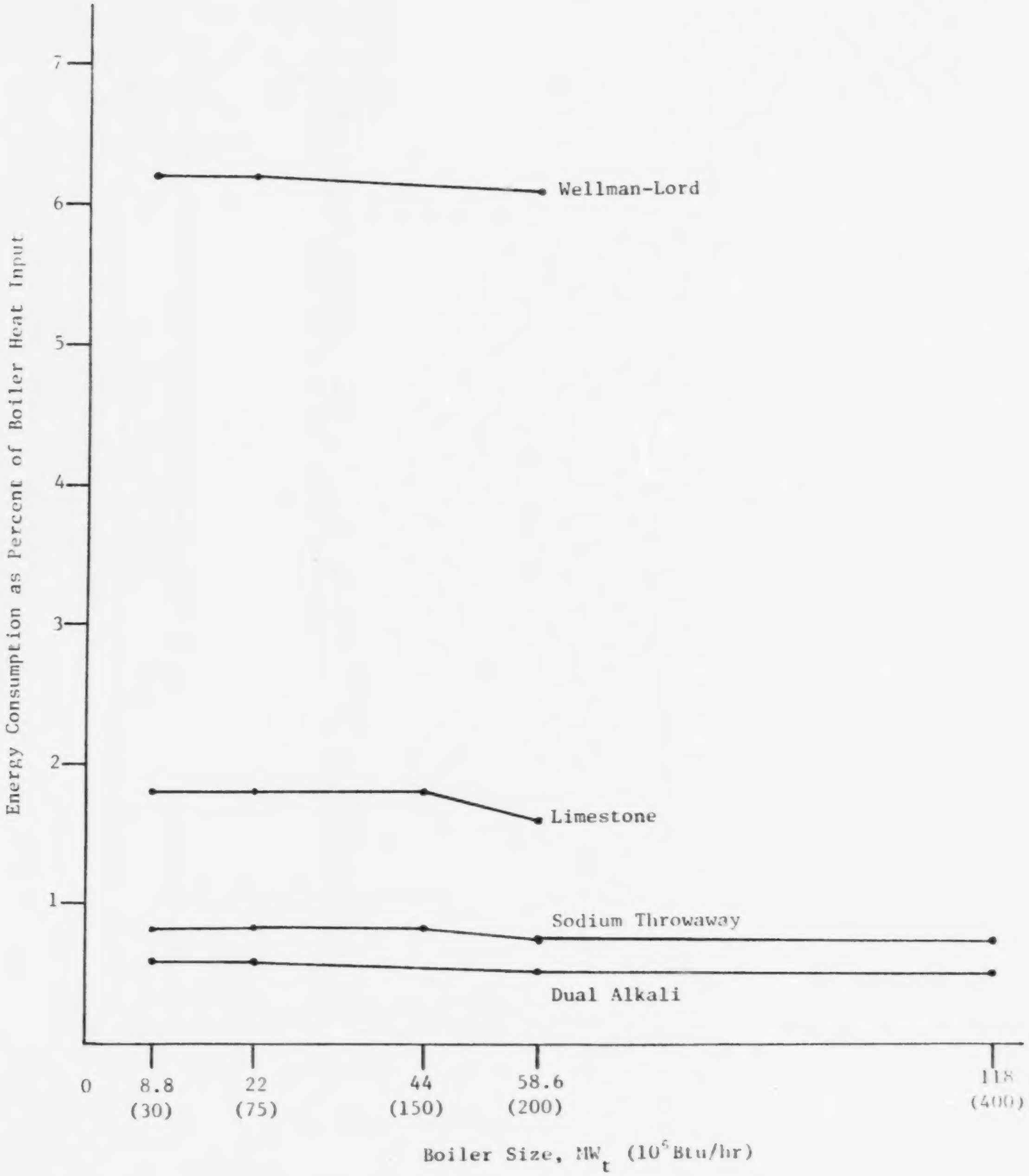


Figure 5.2-4. Percent energy consumption versus boiler size: high sulfur eastern coal

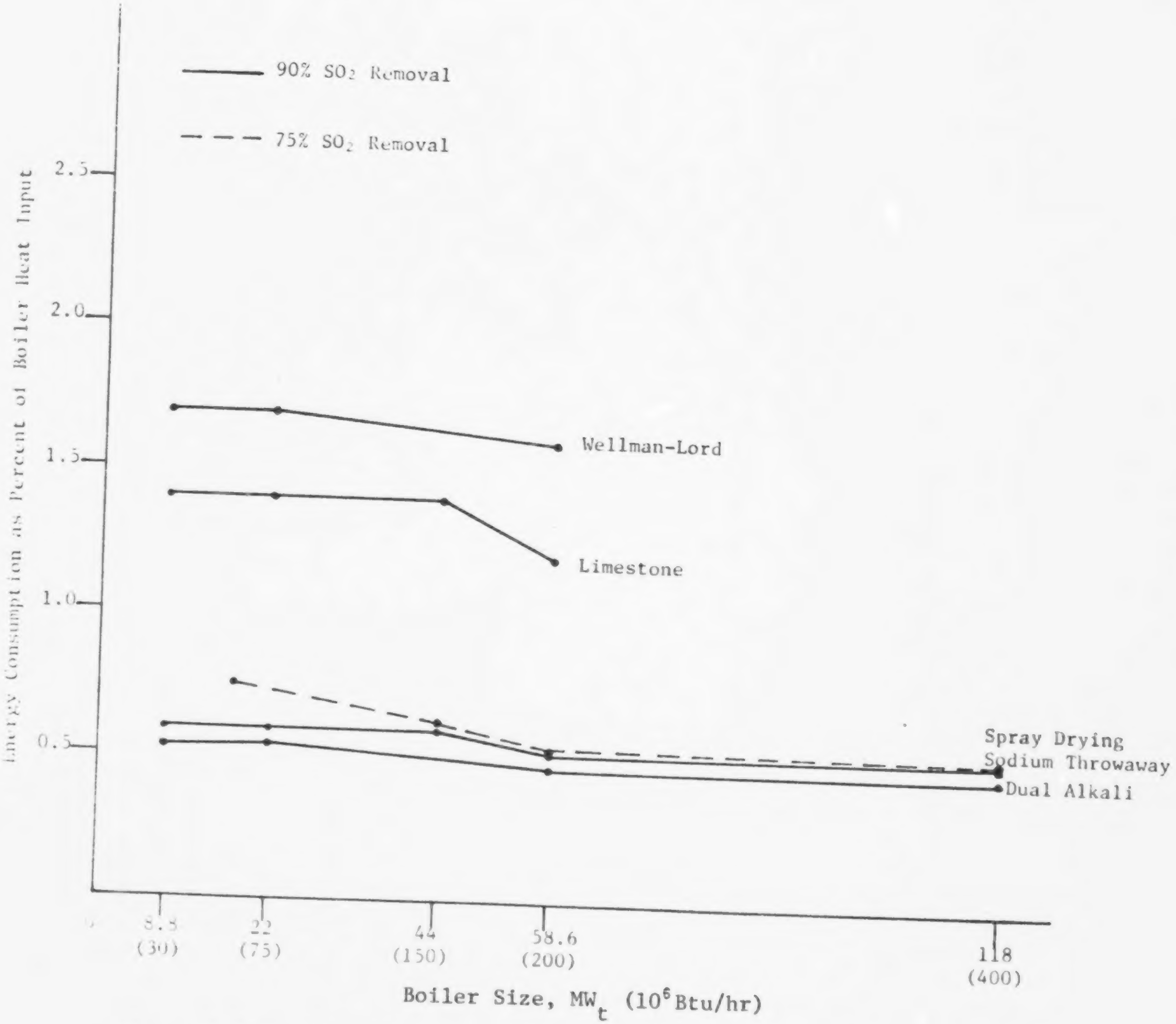


Figure 5.2-5. Percent energy consumption versus boiler size: low sulfur western coal.

Figures 5.2-6 and 5.2-7 illustrate how energy requirements vary with SO_2 removal. These figures show that the energy consumption for the limestone system increases by about 40-50 percent when going from an SO_2 removal of 75 to 90. This is due primarily to an increased liquid pumping rate required to achieve the higher removal efficiency and an increased system pressure drop. Only a very small change in energy consumption (<10 percent) is shown for the sodium throwaway process since its liquid pumping rate is low and does not vary appreciably for a change in SO_2 removal. The change is equally low for the spray drying process as the fan requirements do not change with removal level. Also, the alkali slurry atomization energy requirements did not change appreciably due to the small sized equipment involved. Calculations for the double-alkali and Wellman-Lord processes were only developed for the 90 percent removal cases. Consequently, the effects of varying SO_2 removal levels on process energy requirements are not addressed.

5.2.1 Sample Calculations

The process most widely used in industrial boiler applications is the Sodium Throwaway process; consequently, it was selected for example calculations. The following information was taken from the material balance results presented in Appendix A.

Boiler Size - 58.6 MW (200 10^6 Btu/hr)
Coal Type - 3.5% S Eastern Coal
SO_2 Removal - 90%
Exit Flue Gas Rate - 1430 Nm /min (50,552 scfm)
Liquid Pumping Requirements - 3767 ℓ/min (994 gpm)

From Table 5.2-1:

Discharge Pump Pressure - 3484 Pa (10 psi)
Pumping Height - 6m (20 ft)
Gas Pressure Drop - 100 Pa (8 in. H_2O)

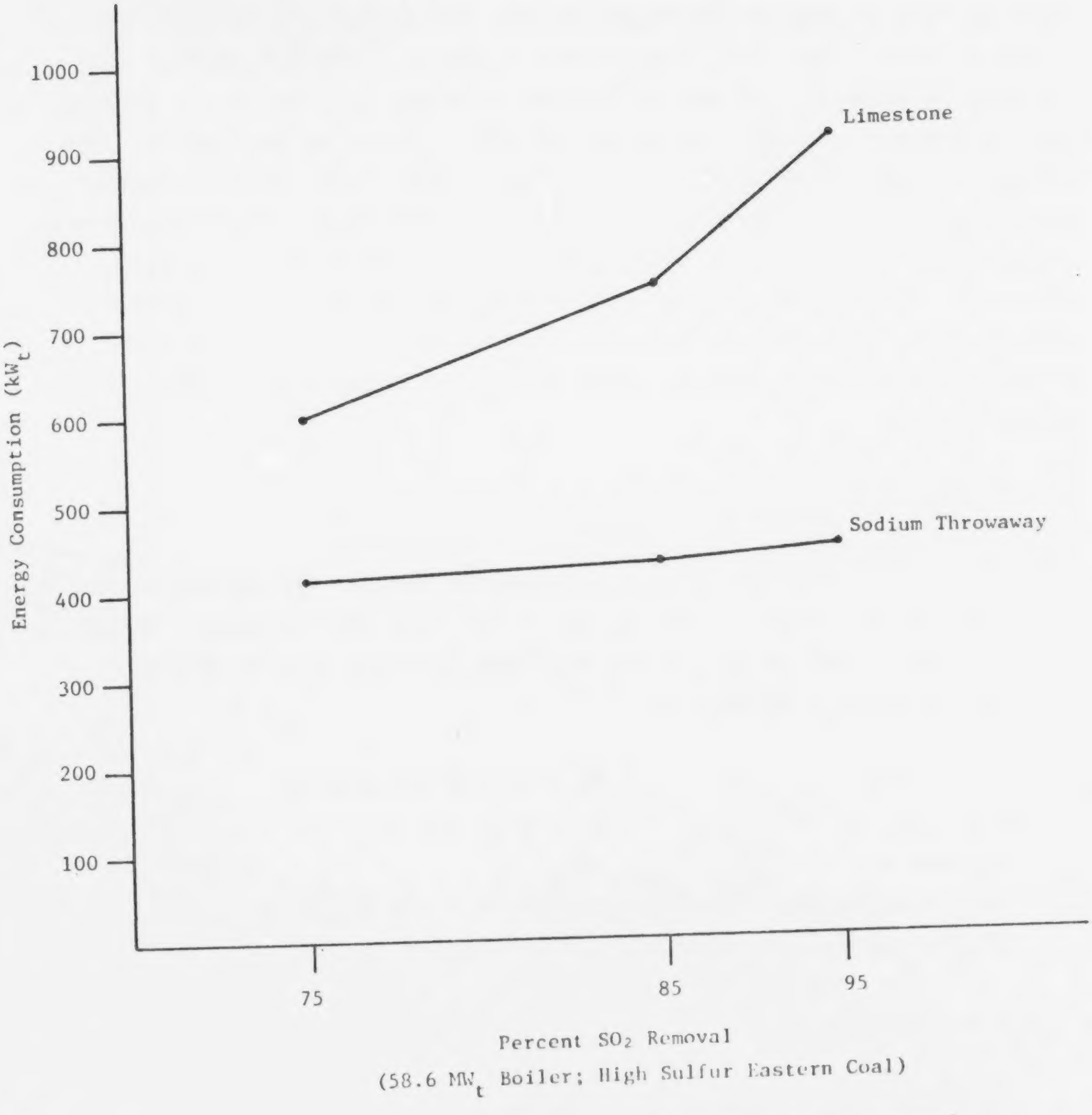
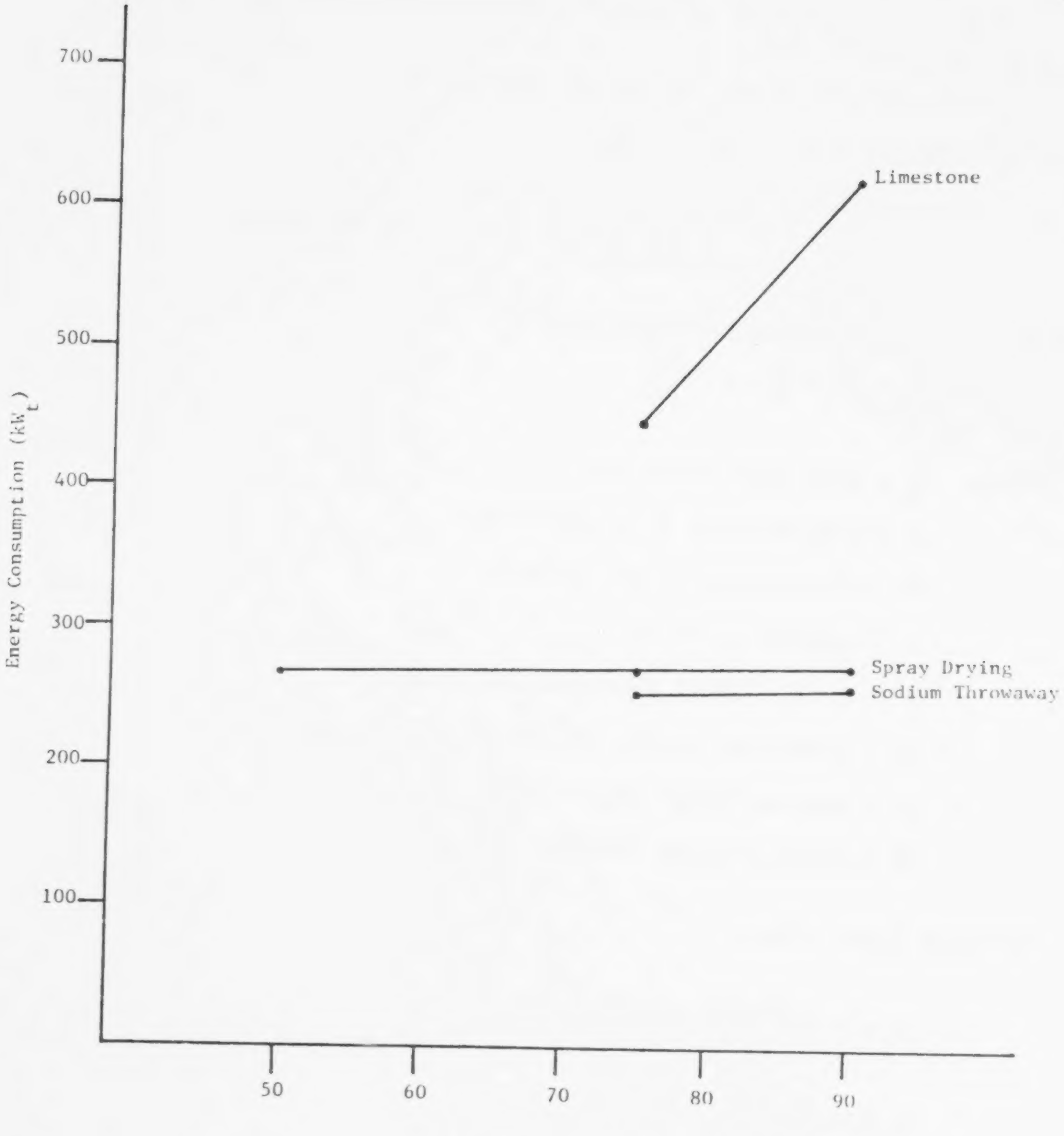


Figure 5.2-6. Energy consumption versus SO$_2$ removal



(44 MW$_t$ Boiler; Low Sulfur Western Coal)

Figure 5.2-7. Energy consumption versus SO_2 removal

5.2.1.1 Calculate Raw Material Handling and Preparation--

From a previous study of energy requirements for soda ash handling and preparation:[11]

Electricity = 24.8 kW/(kg SO_2 removed/sec)

SO_2 removed = 0.127 kg/sec

$$\text{Electricity} = 24.8\ (0.127) = 3.2\ \text{kW}_e$$
$$= 9.3\ \text{kW}_t\ (.032\ 10^6\ \text{Btu/hr})$$

5.2.1.2 Calculate Liquid Pumping Energy--

$$W_p n = \frac{Pb}{\rho} + \frac{g}{gc} Zb + \frac{\bar{V}b^2}{2gc}$$

where: W_p = pump work (ft-lb/sec)

n = pump efficiency [assume 0.6]

Pb = discharge pressure (lb/ft^2)

ρ = density (lb/ft^3)

g = gravitational acceleration (ft/sec^2)

gc = conversion factor (32.179 $ft\text{-}lb_m/lb_f\text{-}sec$)

Zb = pumping height (ft)

$\bar{V}b$ = fluid velocity (ft/sec) [assume 10]

applying these values:

$$W_p n = \frac{(10)(144)}{62.3} + \frac{(32.2)}{32.17}\frac{(20)}{} + \frac{(10)^2}{64.34}$$

$$W_p n = 44.7\ \frac{ft\text{-}lb_f}{lb_m}$$

$$W_p = 74.5\ \frac{ft\text{-}lb_f}{lb_m}$$

To calculate the required power:

$$P_B = \frac{\dot{m}\ W_p}{550}$$

where: P_B = brake horsepower (Hp)

$\dot{m}$ = mass flow rate (lb/sec)

$$\dot{m} = 994\ \frac{\text{gal}}{\text{min}} \times \frac{\text{min}}{60\ \text{sec}} \times \frac{8.33\ \text{lb}}{\text{gal}}$$

$$\dot{m} = 138.0\ \text{lb/sec}$$

$$P_B = \frac{(138.0)(74.5)}{550} = 18.7\ \text{Hp}$$

$$P_B = 18.7\ \text{Hp} \times \frac{\text{kW}}{1.34\ \text{Hp}} = 13.9\ \text{kW}_e$$

$$= 40.7\ \text{kW}_t$$

5.2.1.3 Calculate Fan Energy--

scfm = 50,552 scfm

Pressure drop = 8 in. H_2O

Gas density = .08 lb/ft^3

Fan efficiency = .6

Exit gas velocity = 60 ft/sec

$$W_p n = \frac{\Delta P}{\rho}$$

$$W_p n = \frac{1}{.6}\left[\frac{(8\ \text{in.}H_2O)\ (5.2\ \frac{lb_f/ft^2}{\text{in. }H_2O}}{(.08\ lb_m/ft^3}\right]$$

$$= 867\ \frac{ft\text{-}lb_f}{lb_m}$$

$$P_B = \frac{\dot{m}\ W_p}{(550)(1.34)}$$

$$m = 50{,}522\ \frac{\text{scf}}{\text{min}} \times \frac{\text{mole}}{379\ \text{scf}} \times \frac{29\ \text{lb}}{\text{mole}} \times \frac{\text{min}}{60\ \text{sec}}$$

5.2.1.4 Calculate Waste Disposal Requirements

Assume viscosity of waste water equals that of water -

$\mu = .0012$ lb/ft-sec

Assume linear fluid velocity in pipe -

$v = 10$ ft/sec

Average flow rate for all cases examined:

$$= 39\ \text{GPM}\ \frac{(\text{ft}^3)}{(7.48\ \text{gal.})}\ \frac{(\text{min})}{(60\ \text{sec})}$$

$$= .087\ \text{ft}^3/\text{sec}$$

Cross-sectional area of pipe

$$A = \frac{.087\ \text{ft}^3/\text{sec}}{10\ \text{ft/sec}}$$

$$= .0087\ \text{ft}^2/\text{sec} = \frac{\pi D^2}{4}$$

$$D = .093\ \text{ft}$$

Reynolds Number--

$$Re = \frac{Dv\rho}{\mu}$$

$$= \frac{(.093\ \text{ft})\ (10\ \text{ft/sec})\ (62.3\ \text{lb/ft}^3)}{(.0012\ \text{lb/ft-sec})}$$

$$= 4.84 \times 10^4$$

Friction Factor--

$$f = .0052$$

Assume wastewater is pumped to onsite treatment facility, one mile from FGD system--

$$L = 5280\ \text{ft}$$

Friction Loss--

$$F = \left(\frac{4\ fL}{D}\right)\ \left(\frac{V^2}{2gc}\right)$$

$$= \frac{9(.0052)\ (5280\ ft)}{(.093\ ft)}\ \ \frac{(10\ ft/sec)^2}{(2)\ (32.2\ \frac{lb_m\ ft}{sec^2 lb_f})}$$

$$= 1834\ \frac{ft\text{-}lb_f}{lb_m}$$

Pump Work--

$$W_p n = \frac{P_b}{\rho} + \frac{g}{gc} Z_b + \frac{\bar{V}_b{}^2}{2\ gc} + F$$

$$= \frac{10\ (144)}{62.3} + 0 + \frac{(10)^2}{2(32.2)} + 1834$$

$$= 23 + 0 + 2 + 1834$$

$$= 1859\ \frac{ft\text{-}lb_f}{lb_m}$$

$$Wp = \frac{1859}{.6} = 3100\ \frac{ft\text{-}lb_f}{lb_m}$$

Required Power--

$$P_B = \frac{\dot{m}\ W_p}{550}$$

from Appendix A--

flow = 92 gpm

$$\dot{m} = 92\ \frac{gal}{min}\ X\ \frac{min}{60\ sec}\ X\ \frac{8.33\ lb}{gal}$$

$$= 13\ lb_m/sec$$

$$P_B = \frac{(13\ lb_m/sec)\ (3100\ ft\text{-}lb_f/lb_m)}{550\ \frac{ft\text{-}lb_f/sec}{hp}}$$

$$= 73.3 \text{ hp } \left(\frac{\text{KW}}{1.34 \text{ hp}}\right)$$

$$= 54.7 \text{ kW}_e$$

$$= 160.4 \text{ kW}_t$$

5.2.1.5 Calculate Utilities and Services--

From a previous study of energy requirements:[12]

Electricity = 0.15 kW/(Nm^3/sec)

Inlet flow rate = 45,228 scfm

$$\frac{45{,}228 \text{ ft}^3}{\text{min}} \left| \frac{492°\text{R}}{520°\text{R}} \right| \frac{1 \text{ min}}{60\text{s}} \left| \frac{(.3048)^3 \text{m}^3}{1 \text{ ft}^3} \right. = 21 \text{ Nm}^3/\text{s}$$

$$\text{Electricity} = .15(21) = .2 \text{ kW}_e$$

$$= 9.3 \text{ kW}_t \ (.033\ 10^6 \text{ Btu/hr})$$

5.2.1.6 Calculation Summary--

	kW_t	10^6 Btu/hr
Raw materials preparation/handling	9.3	0.03
Liquid pump energy requirements	40.7	0.14
Fan energy requirements	222.0	0.76
Wastewater disposal pumping	160.4	0.55
Utilities and services	9.3	0.03
	441.7	1.51

5.2.2 Methods to Reduce Energy Consumption

Energy consumption in nonregenerable FGD systems that do not use reheat is primarily attributed to two main sources: electricity for driving fans and electricity for driving pumps. Other significant energy consumers are the feed preparation and utilities and services. For small industrial boiler applications electrical power requirements will probably be purchased whereas steam requirements may result in boiler derating. Both electricity and steam may have to be purchased for some installations such as those for boilers producing hot water only. Table 5.2-7 presents a summary of the relative percentage of each of the energy requirements of each process area as compared to the overall energy requirement for the FGD process. This table shows that fans are the largest energy consumer for each of the wet throwaway processes when stack gas reheat energy requirements are not included. However, when reheat energy requirements are included, they become the dominant energy consuming portion of these processes.

TABLE 5.2-7. PERCENTAGE ENERGY CONSUMPTIONS FOR NONREGENERABLE PROCESSES (58.6 MW_t Boiler, 90% Removal, Eastern Coal)

Source of energy consumption	Limestone		Double Alkali		Sodium TA	
	kW_t	Percent of total	kW_t	Percent of total	kW_t	Percent of total
Raw materials handling and preparation	114.2	13	12.7	1	9.3	2
Liquid pumps	317.4	35	63.2	21	40.7	9
Fans	471.7	51	222.0	72	222.0	51
Disposal pumps	-	-	-	-	160.4	36
Utilities and services	8.6	1	9.3	3	9.3	2
TOTAL	911.9		307.2		441.7	
Reheat Steam	884.0		884.0		884.0	
	1795.9		1191.2		1325.7	

Liquid pumping requirements are set by the required SO_2 removal; thus it is unlikely that they will be reduced for a given process design. However, changes in a process design such as using a different sorbent (for example lime *vs.* limestone), a different gas-liquid contactor, or using additives to enhance the liquid phase alkalinity may all result in reduced liquid pumping rates. Fan energy requirements are a function of the system pressure drop and will vary with the type of gas-liquid contactor and the volume of flue gas entering the system. Use of an open type contactor (for example a spray tower) coupled with a decrease in the amount of excess air used by the boiler will result in decreases in the fan energy requirements.

Energy requirements for the spray drying process are already quite low, and compare well with the other nonregenerable processes when no reheat is included. Table 5.2-8 illustrates this for a low sulfur western coal case. If reheat is required, the spray drying process has a significant advantage since it exhausts its flue gas at 175°F and requires no reheat.

TABLE 5.2-8. COMPARISON OF SPRAY DRYING PROCESS ENERGY REQUIREMENTS
(44 MW_t, 0.6% S coal, 75% Removal)

Source of energy consumption	Limestone		Sodium TA		Lime Spray Drying	
	kW_t	Percent of total	kW_t	Percent of total	kW_t	Percent of total
Raw Material Handling	15	3	1	-	1.5	1
Liquid Pumping	133	30	29	11	0.5	-
Fans	292	65	195	78	140	52
Disposal Pumps	-	-	21	8	-	-
Atomization	-	-	-	-	120	44
Utilities & Services	8	2	9	3	8	3
Total	448		254		270	100
Reheat steam	775		775		-	
	1223		1029		270	100

Excluding reheat, energy consumption in the Wellman-Lord process is attributable to four main sources: electricity for pumps and fans, process steam, and methane for reducing SO_2 to sulfur. Table 5.2-9 shows the relative energy requirements of each of the Wellman-Lord process areas for both eastern and western coal applications.

For this process, the relative amounts of energy consumed by the various process areas varies depending upon the sulfur content of the coal being burned. However, for both the eastern 3.5 percent sulfur coal case and the western 0.6 percent sulfur coal cases, the regeneration and sulfur production areas are the major energy users. It is doubtful that the energy requirements of the regeneration processing area can be reduced since double effect evaporators were assumed for these calculations which are some 45 percent more energy efficient than single effect evaporators.[13] If reheat energy requirements are included, they become the major energy consumer for the low-sulfur western coal case and serve to increase the energy requirements of both cases by about 900 kW_t.

TABLE 5.2-9. PERCENTAGE ENERGY CONSUMPTION FOR WELLMAN-LORD PROCESS
($58.6\ MW_t$ 90% SO_2 Removal)

Source of energy consumption	Western coal (0.6%S)		Eastern coal (3.5%S)	
	kW_t	Percent of total	kW_t	Percent of total
Raw materials handling and preparation	1.9	<1	9.3	<1
Pumps	42.2	4	82.9	2
Fans	208.6	22	205.5	6
Process steam	469.0	49	2220.0	62
Methane	219.0	23	1048.0	29
Utilities and services	10.2	1	9.6	<1
Total	950.9		3575.3	
Reheat steam	900.0	48.6	884.0	20
	1850.9		4459.3	

Energy requirements in the SO_2 reduction area could be decreased if sulfuric acid were produced rather than sulfur. Acid production does not require a reducing gas. Calculations to compare the relative energy consumption of sulfuric acid versus sulfur production were not performed as part of this study.

5.3 IMPACT OF CONTROLS FOR OIL-FIRED BOILERS

The energy impact of FGD systems applied to oil-fired boilers will be similar to that for coal-fired boilers. The major difference will be that the oil-fired boilers will generally operate with a lower percent excess air which will result in a lower energy requirement for the fans. The standard residual oil-fired boiler operates with only 15 percent excess air whereas the standard coal-fired boiler of the same size, 44 MW (150×10^6 Btu/hr) operates with 50 percent excess air.

Table 5.3-1 presents the results of calculations to estimate the energy requirements for a limestone system applied to a residual oil-fired boiler. This table also presents the total energy requirements for the limestone system applied to a 44 MW (150×10^6 Btu/hr) coal-fired boiler. The system applied to the residual oil-fired boiler has energy requirements on the order of 70 percent of those for the systems treating flue gases from coal-fired boilers. By analogy, energy requirements for other nonregenerable or throw-away processes should be on the order of 70 percent of those for coal-fired applications.

For the Wellman-Lord process, a significant amount of the overall process energy consumption occurs in the regeneration and SO_2 conversion sections of the process. Because of this, energy consumption for a residual oil application will vary depending upon the amount of SO_2 processed. As shown in Figure 5.2-1, the energy consumption of a Wellman-Lord system is strongly related to the sulfur content of the fuel being burned. From Figures 5.2-2 and 5.2-3, for a 44 MW (150×10^6 Btu/hr) application, a Wellman-Lord system applied to an eastern 3.5%S coal-fired boiler would consume about 2700 kW whereas a Wellman-Lord system applied to a western 0.6% coal-fired boiler would consume about 700 kW. The eastern coal application would process

TABLE 5.3-1. LIMESTONE PROCESS ENERGY REQUIREMENTS FOR RESIDUAL OIL APPLICATION
[44 MW_t(150 x 10^6 Btu/hr)]

Source of energy consumption	Removal levels, percent					
	90		85		75	
	kW	% of total	kW	% of total	kW	% of total
Raw materials handling and preparation	47.7	9.8	45.7	11.3	39.8	12.0
Pumps	172.2	35.2	135.5	33.5	99.1	30.0
Fans	263.5	53.8	217.0	53.8	186.0	56.3
Disposal	1.0	.2	.9	.2	.8	.2
Utilities and Services	4.8	1.0	4.8	1.2	4.8	1.5
Total	489.2		403.9		330.5	
Percent of boiler heat input	1.1		0.92		0.75	
Total kW for a limestone system applied to a 44 MW boiler burning 3.5% eastern coal	776.8		—		507.3	
Total kW for a limestone system applied to a 44 MW boiler burning 0.6% S western coal	620.7		—		447.7	

15.8 moles/hr SO_2 whereas the western coal application would process 3.7 moles/hr SO_2.

Energy requirements for a Wellman-Lord system applied to the standard residual oil-fired boiler are shown in Table 5.3-2. Energy requirements for this system for the 90 percent SO_2 removal case (removal of 6.6 moles/hr SO_2) were calculated to be 1687 kW. This value compares well with the energy requirements for the coal-fired Wellman-Lord cases as discussed above.

Previous comments for coal-fired FGD system applications concerning major energy consumption sources and potential methods of energy reduction will apply to residual oil-fired FGD systems. Also, all of the energy impacts discussed in this section will apply to both new and retrofit applications.

TABLE 5.3-2. WELLMAN-LORD PROCESS ENERGY REQUIREMENTS FOR RESIDUAL OIL APPLICATION
[44 MW_t(150 x 10^6 Btu/hr)]

Source of energy consumption	Removal levels, percent					
	90		85		75	
	kW	% of total	kW	% of total	kW	% of total
Raw materials handling and preparation	3.9	0.2	3.6	0.2	3.2	0.2
Pumps	54.4	3.2	53.9	3.3	53.3	3.6
Fans	114.5	6.8	114.5	7.1	114.5	7.7
Process Steam	1064.0	63.1	1016.0	62.6	886.0	60.0
Methane	445.0	26.4	430.0	26.5	416.0	28.2
Utilities and services	4.8	0.3	4.8	0.3	4.8	0.3
Total	1686.6		1622.8		1477.8	
Percent of boiler heat input	3.8		3.7		3.4	

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SECTION 6

ENVIRONMENTAL IMPACT OF CANDIDATES FOR BEST EMISSION CONTROL SYSTEMS

6.0 INTRODUCTION

Each of the flue gas desulfurization (FGD) processes selected as candidates for industrial boiler application create beneficial air emission impacts by reducing both the SO_2 and particulate emissions from industrial boilers. However, adverse water and solid waste emission impacts can result from FGD systems if proper design and operating practices are not followed since the SO_2 removed from the boiler flue gas is converted to liquid or solid by-products. The limestone and double alkali processes convert the absorbed SO_2 into sludges; the spray drying process converts the absorbed SO_2 into a dry solid; and the sodium throwaway process converts the absorbed SO_2 into soluble sodium salts. Although the Wellman-Lord process produces a salable by-product, either sulfur or sulfuric acid, a purge sodium sulfate solids stream and a high chloride aqueous stream are produced which must be properly disposed of. This chapter examines the air water, and solid waste impacts of each of the candidate FGD processes.

6.2 ENVIRONMENTAL IMPACT OF CONTROLS FOR COAL-FIRED BOILERS

6.2.1 Air Pollution

Four of the five FGD processes under consideration are wet scrubbing systems and as such have the capability of removing particulates as well as

SO_2 from industrial boiler flue gases. In fact, several industrial boiler FGD systems in use today are designed and operated in a dual particulate/SO_2 removal service. The fifth candidate process, spray drying, is a dry process with down stream particulate collection (baghouse or ESP) to collect the solid waste product, generated in the spray dryer along with any accompanying particulate material.

Although the FGD systems under consideration have the potential for combined SO_2/particulate removal, only the SO_2 removal aspects of these systems will be considered in detail in this Individual Technology Assessment Report (ITAR). Particulate removal is considered in detail in the particulate collection technology assessment report.

Particulate removal in FGD systems has been evaluated by EPA as part of their efforts to prepare the proposed New Source Performance Standards reported in the 19 September Federal Register for Electric Utility Steam Generating Units.[1] As reported in the 19 September 1978 Federal Register, the EPA has investigated the following mechanisms to attempt to assess the impacts of FGD systems upon particulate emissions:

1) FGD system sulfate carryover from the scrubber slurry,

2) particulate matter removal by the FGD systems, and

3) particulate matter generation by the FGD system through condensation of sulfuric acid (H_2SO_4) mist.

With regard to the first mechanism, data were taken from steam generators that had operating FGD systems with low particulate levels at the FGD inlet. The data indicated that with a properly maintained mist eliminator, particulate emissions did not increase through the FGD system.[2] With regard to the second mechanism, the FGD system data indicated that scrubbers did indeed reduce the level of particulate emissions.[3]

With regard to the third mechanism, the EPA obtained data that indicated an FGD system applied to a low sulfur coal-fired boiler will not increase particulate emissions through sulfuric acid formation. Sufficient data, however, have not become available for high sulfur coal-fired applications to fully assess the effect of sulfuric acid emissions on measured particulate emissions and this matter is currently undergoing further investigation by EPA.[4]

With regard to SO_2, the air pollution impact of the FGD processes will be the same for a given SO_2 removal level. Tables 6.2-1 through 6.2-5 show the emission levels achieved by applying varying SO_2 control levels to the standard coal-fired boilers. Figures 6.2-1 through 6.2-3 illustrate these impacts graphically. As shown in Tables 6.2-1 through 6.2-5, emissions from the 3.5 percent sulfur eastern coal fired boilers will be about 4.7 times those from the 0.6 percent sulfur western coal-fired boilers and 1.7 times those from the 2.3 percent sulfur coal-fired boilers for an equivalent level of control. This is illustrated in the three figures mentioned previously and, in Figure 6.2-4.

The amount of SO_2 removal necessary to achieve compliances with an average State Implementation Plan (SIP) control level was also considered. The average SIP control level was defined by EPA for this study to be 1075 ng/J (2.5 lb/10^6 Btu). It is interesting to note that no SO_2 control would be required for a boiler firing 0.6 percent sulfur western coal to comply with this level whereas a boiler burning 3.5 percent sulfur eastern coal would require an SO_2 control level of 56 percent to meet this same level and a boiler burning 2.3 percent sulfur coal would require 24 percent SO_2 control.

A secondary type of environmental impact resulting from the application of FGD systems to industrial boilers are problems associated with the formation of a saturated flue gas plume in the gas-liquid contactor. Stack gases from industrial boilers without FGD are normally exhausted at temperatures of between 150-200°C (300-400°F). In this temperature range, the flue gas is relatively dry and noncorrosive. The wet FGD processes cool the boiler flue gas to its adiabatic saturation temperature, normally 50-60°C (125-140°F). This saturated flue gas may cause the followed problems:

TABLE 6.2-1. AIR POLLUTION IMPACTS OF SO_2 CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

Boiler size and type	Control level (%)	SO_2 Emissions			
		11,800 Btu/lb 3.5% S eastern coal		9,600 Btu/lb 0.6% S western coal	
		g/s (lb/hr)	ng/J (lb/10^6 Btu)	g/s (lb/hr)	ng/J (lb/10^6 Btu)
8.8 MW_t (30 X 10^6 Btu/hr) Underfeed stoker	Uncontrolled	21.3 (169)	2408 (5.6)	4.5 (35.8)	516 (1.2)
	SIP*	9.5 (75)	1075 (2.5)		
	Moderate 75	5.3 (42.2)	606 (1.4)	1.1 (9.0)	128.9 (.3)
	Intermediate 85	3.2 (25.6)	365 (.85)	.7 (5.4)	77.0 (.2)
	Stringent 90	2.1 (16.6)	236.5 (.6)	.5 (3.6)	51.6 (.1)

*56 percent removal required to meet average SIP for 3.5% S eastern coal, no FGD required for 0.6% S western coal.

TABLE 6.2-2. AIR POLLUTION IMPACTS OF SO_2 CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

Boiler and type	Control level (%)	SO_2 emissions					
		11,800 Btu/lb 3.5% S eastern coal		13,200 Btu/16 2.3% S coal		9,000 Btu/lb 0.6% S western coal	
		g/s (lb/hr)	ng/J ($lb/10^6$ Btu)	g/s (lb/hr)	ng/J ($lb/10^6$ Btu)	g/s (lb/hr)	ng/J ($lb/10^6$ Btu)
22 MW_t ($75 X 10^6$ Btu/hr) Chaingrate Stoker	Uncontrolled	53.3 (423)	2408 (5.6)	31.3 (248.3)	1421 (3.3)	11.2 (88.9)	516 (1.2)
	SIP*	23.6 (187.5)	1075 (2.5)	23.6 (187.5)	1075 (2.5)		
	Moderate 75	13.3 (105.6)	605.3 (1.4)	7.8 (62.1)	355.3 (.83)	2.8 (22.4)	128.9 (.3)
	Intermediate 85	8.0 (63.4)	361.1 (.8)	4.7 (37.2)	213.2 (.50)	1.8 (13.3)	77.4 (.2)
	Stringent 90	5.3 (42.2)	242.1 (.6)	3.1 (24.8)	142.1 (.33)	1.1 (9.0)	51.6 (.1)

*56 percent removal required to meet average SIP for 3.5% S eastern coal, 24 percent removal for 2.3% S coal, and no FGD required for 0.6% S western coal.

TABLE 6.2-3. AIR POLLUTION IMPACTS FROM "BEST" SO_2 CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

Boiler size and type	Control level (%)	SO_2 Emissions			
		11,800 Btu/lb 3.5% S eastern coal		9,600 Btu/lb 0.6% S western coal	
		g/s (lb/hr)	ng/J (lb/10^6 Btu)	g/s (lb/hr)	ng/J (lb/10^6 Btu)
44 MW_t (150 X 10^6Btu/hr) Spreader Stoker	Uncontrolled	106.4 (844.8)	2408 (5.6)	22.4 (177.9)	516 (1.2)
	SIP*	47.3 (375)	1075 (2.5)		
	Moderate 75	26.6 (211.2)	605.3 (1.4)	5.6 (44.2)	126.6 (.3)
	Intermediate 85	16.0 (176.7)	361.1 (.8)	3.4 (26.7)	77.4 (0.2)
	Stringent 90	10.6 (84.5)	242.1 (.6)	2.3 (17.9)	51.4 (.1)

*56 percent removal required to meet average SIP for 3.5% S eastern coal, no FGD required for 0.6% S western coal.

TABLE 6.2-4. AIR POLLUTION IMPACTS FROM "BEST" SO_2 CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

		SO_2 Emissions			
		11,800 Btu/lb 3.5% S eastern coal		9,600 Btu/lb 0.6% S western coal	
Boiler size and type	Control level (%)	g/s (lb/hr)	ng/J (lb/10^6 Btu)	g/s (lb/hr)	ng/J (lb/10^6 Btu)
58.6 MW_t (200 X 10^6 Btu/hr) Pulverized Coal	Uncontrolled	141.9 (1126.4)	2408 (5.6)	29.9 (237.4)	516 (1.2)
	SIP*	63.0 (500)	1075 (2.5)		
	Moderate 75	35.5 (281.6)	605.3 (1.4)	7.5 (59.5)	128.9 (.3)
	Intermediate 85	21.3 (169)	363.2 (.8)	4.5 (35.8)	77.0 (.2)
	Stringent 90	14.2 (112.6)	242.1 (.6)	3.0 (23.7)	51.4 (.1)

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*56 percent removal required to meet average SIP for 3.5% S eastern coal, no FGD required for 0.6% S western coal.

TABLE 6.2-5. AIR POLLUTION IMPACTS FROM "BEST" SC$_2$ CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

Boiler and type	Control level (%)	SO_2 emissions 11,800 Btu/lb 3.5% S eastern coal g/s (lb/hr)	 11,800 Btu/lb 3.5% S eastern coal ng/J (lb/10^6 Btu)	 13,200 Btu/16 2.3% S coal g/s (lb/hr)	 13,200 Btu/16 2.3% S coal ng/J (lb/10^6 Btu)	 9,000 Btu/lb 0.6% S western coal g/s (lb/hr)	 9,000 Btu/lb 0.6% S western coal ng/J (lb/10^6 Btu)
118 MW_t (400 X 10^6 Btu/hr) Pulverized Coal	Uncontrolled	284.3 (2254.4)	2423 (5.6)	166.9 (1323.2)	1422 (3.3)	59.9 (474.9)	510.5 (1.2)
	SIP*	126.1 (1000.0)	1075 (2.5)	126.1 (1000.0)	1075 (2.5)		
	Moderate	71.1 (563.6)	605.9 (1.4)	41.7 (330.8)	335.6 (0.8)	15.0 (118.7)	127.6 (0.3)
	Intermediate 85	42.7 (338.2)	363.6 (0.8)	25.0 (198.5)	213.4 (0.5)	9.0 (71.2)	76.5 (0.2)
	Stringent 90	28.4 (225.4)	242.3 (0.6)	16.7 (132.3)	142.2 (0.3)	6.0 (47.5)	51.1 (0.1)

*56 percent removal required to meet average SIP for 3.5% S eastern coal, 24 percent removal for 2.3% S coal, and no FGD required for 0.6% S western coal.

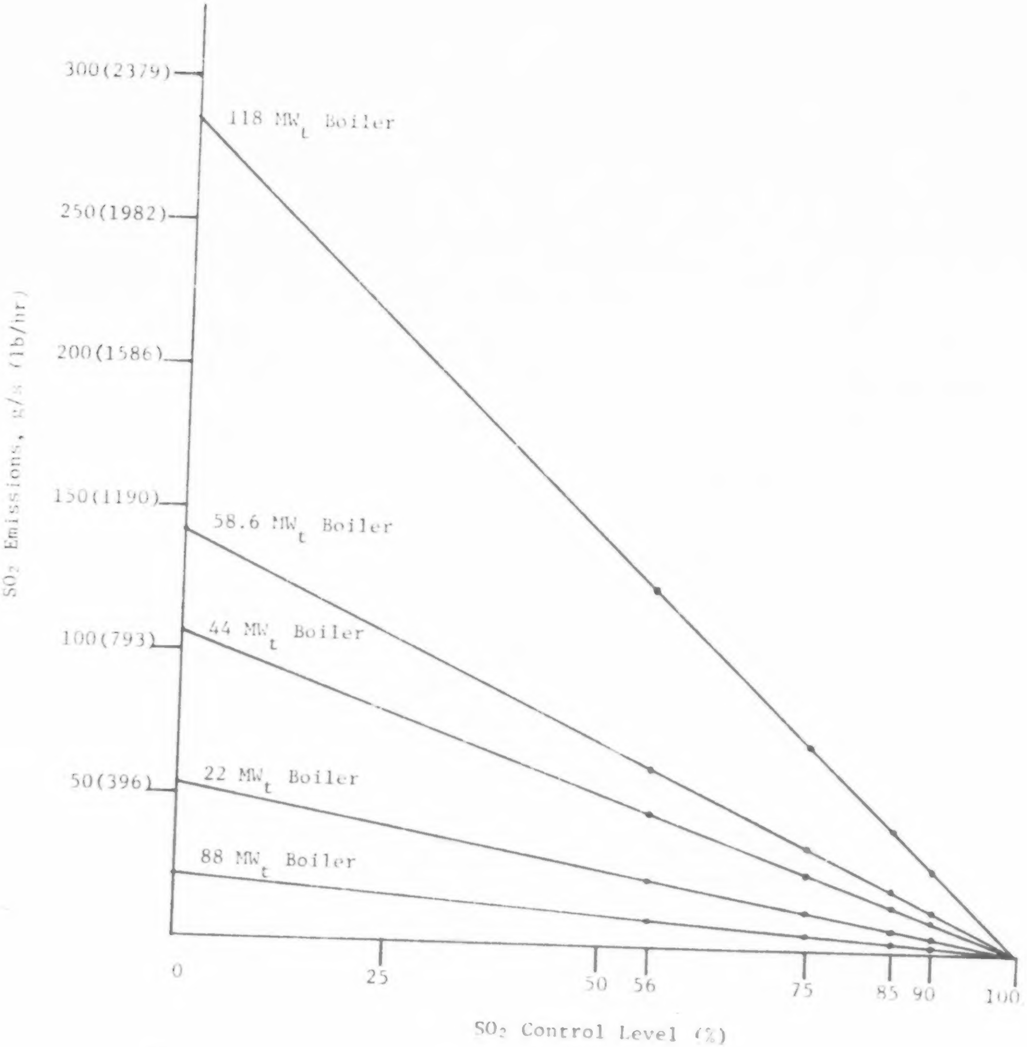

Figure 6.2-1. SO_2 EMISSIONS VERSUS CONTROL LEVEL FOR STANDARD BOILERS FIRING 3.5% SULFUR EASTERN COAL

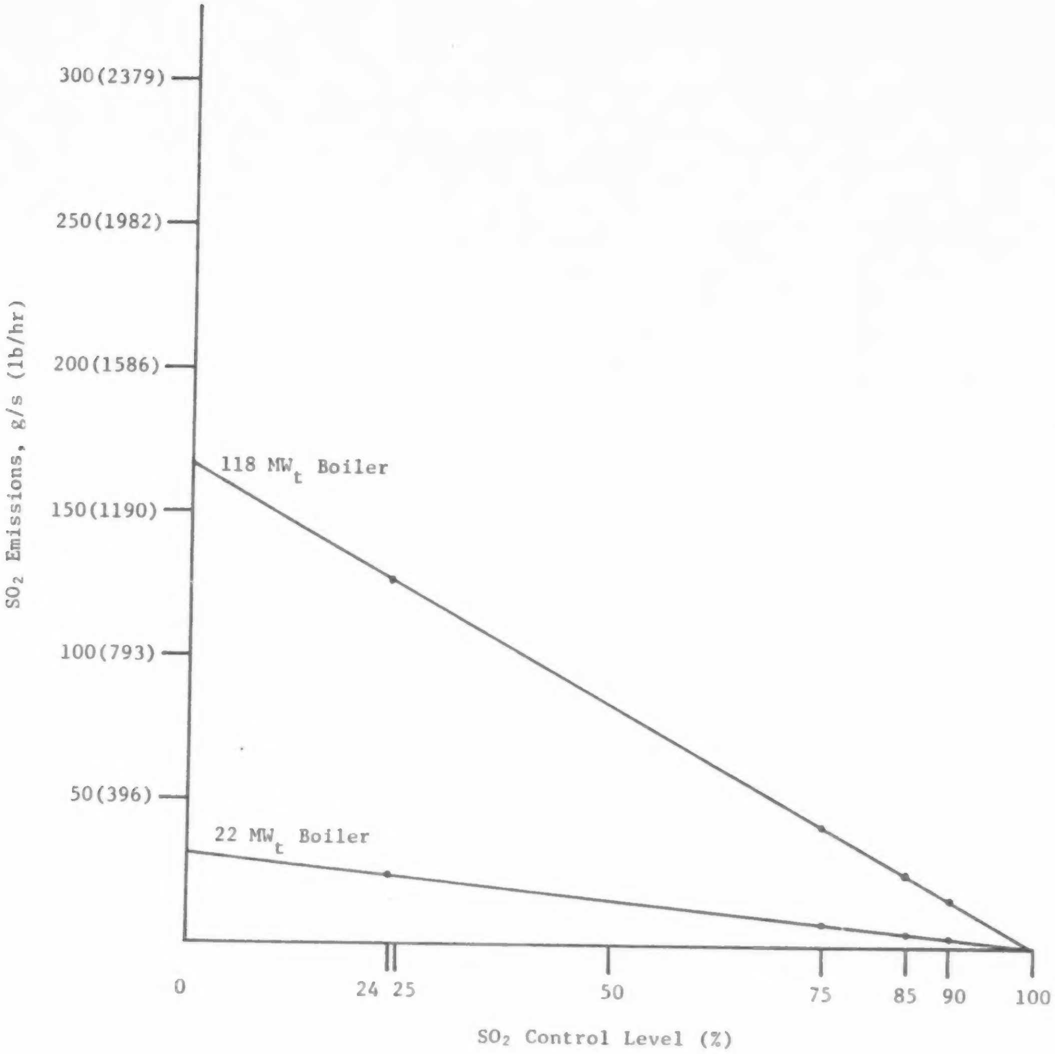

Figure 6.2-2. SO_2 EMISSIONS VERSUS CONTROL LEVEL FOR STANDARD BOILERS FIRING 2.3% SULFUR COAL.

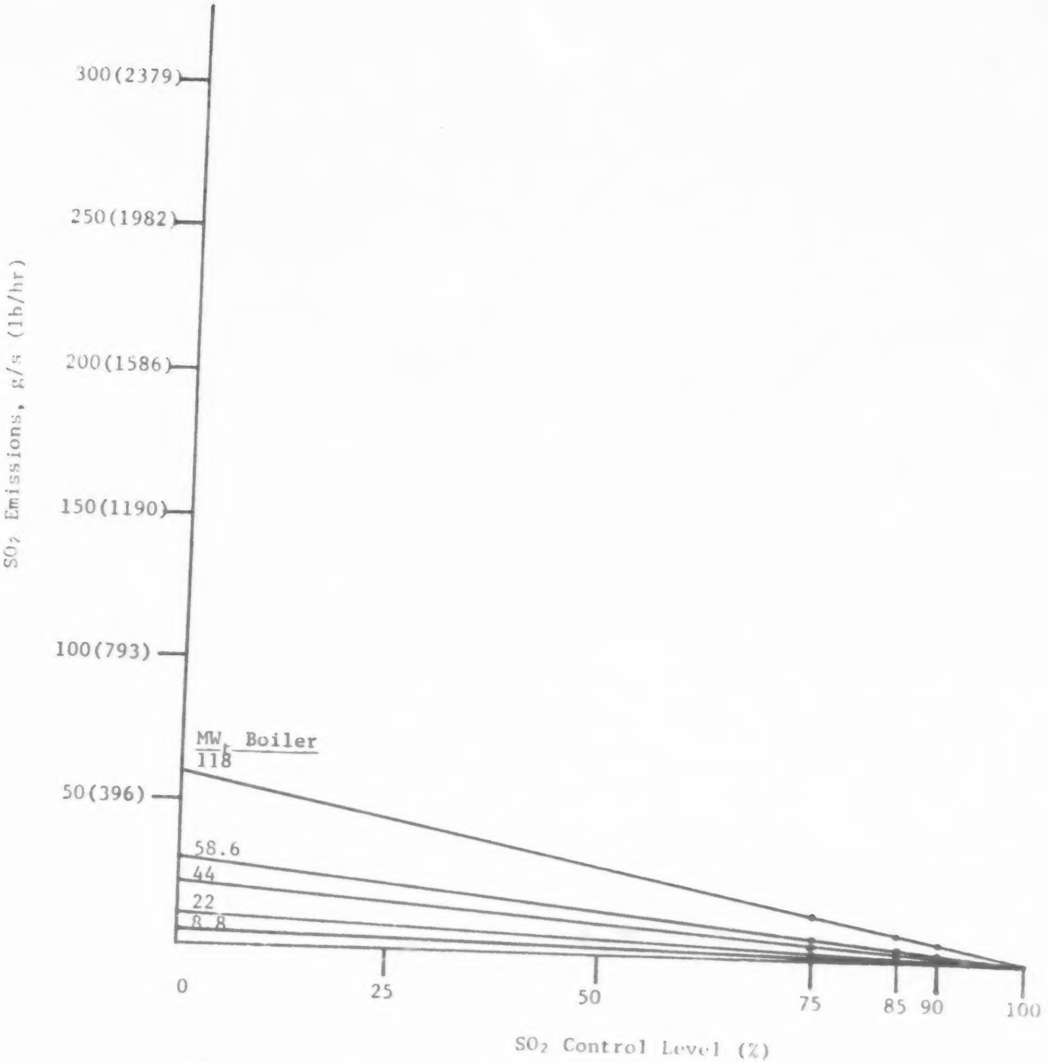

Figure 6.2-3. SO_2 EMISSIONS VERSUS CONTROL LEVEL FOR STANDARD BOILERS FIRING 0.6% SULFUR WESTERN COAL.

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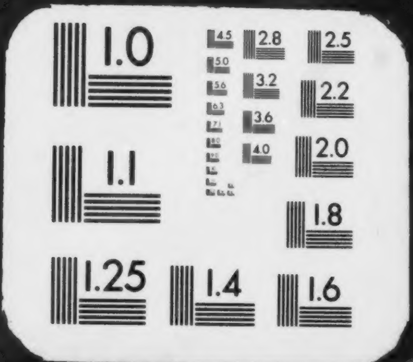

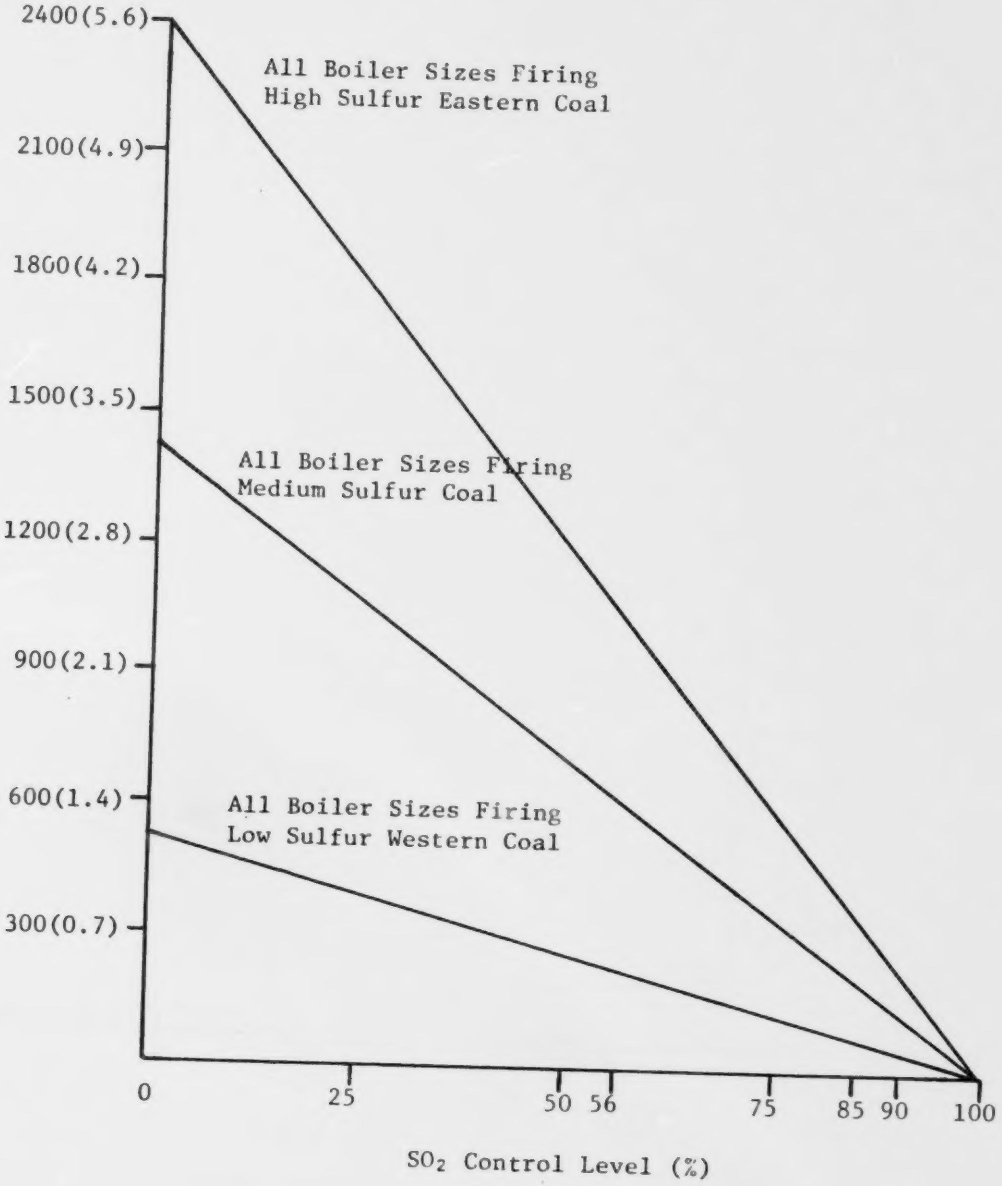

Figure 6.2-4. SO_2 emissions versus control level.

- The occurrence of acid rain in the vicinity of the plant's stack,
- High ground level pollutant concentrations downwind from the stack due to poor plume buoyancy,
- The formation of a visible plume which may be hazardous to any ground and air traffic in the vicinity of the power plant, and
- The corrosion of equipment downstream of the scrubber due to the presence of moisture, acid, and chlorides.

Reheating the saturated flue gas to a temperature above its saturation temperature will lessen the impacts of each of these four potential problems. EPA is currently investigating the amount of reheat required and the best method of achieving the desired amount of reheat to prevent these potential problems.[5]

6.2.2 Water Pollution

Four of the five candidate processes produce waste products that have the possibility of resulting in aqueous emissions. Table 6.2-6 gives the major physical properties of these waste product streams.

TABLE 6.2-6. PHYSICAL PROPERTIES OF WASTE PRODUCTS

System	Waste product physical properties
Sodium Throwaway	Aqueous stream - 5% dissolved solids
Limestone	Sludge - 35-60% solids
Double-Alkali	Sludge - 35-60% solids
Wellman-Lord	Purge solids Aqueous stream - low pH from prescrubber

Good design and operating practices for the limestone and double-alkali processes include dewatering the sludge and recycling the supernatant liquid. Consequently, there should not be water emissions from these systems except for times of severe rainfall or process upsets. Waste streams from these processes will be discussed further in the section on solid waste.

The aqueous waste stream from the prescrubber of the Wellman-Lord process will be characterized by a low pH which results from the chlorides that are removed from the gas stream. In addition, residual fly ash not removed by the upstream particulate removal device may be present in the stream. Table 6.2-7, which gives results of the material balance calculation presented in Appendix A, shows estimated prescrubber discharge rates for the standard size boilers.

TABLE 6.2-7. WELLMAN-LORD PRESCRUBBER DISCHARGES

Boiler size	Discharge rate		
	ℓ/min	gal/min	gal/10^6 Btu heat input
8.8 MW_t	9.1	2.4	4.8
22 MW_t	22.7	6.0	4.8
44 MW_t	45.4	12.0	4.8
58.6 MW_t	60.6	16.0	4.8

This prescrubber discharge stream is for the purpose of maintaining suspended and dissolved solids at desired levels. Suspended solids (fly ash) are generally held at less than five percent whereas the dissolved solid level may be greater than 20,000 mg/ℓ depending upon the chlorine concentration in the coal.

Except for the high chloride concentrations and low pH, the quality of the prescrubber discharge will be very similar to that of the boiler ash sluice water.[6] Since this stream has been estimated to be approximately one percent of the ash sluicing requirements for a power plant, it can be used for ash sluicing where it will become diluted and neutralized with the other ash sluice water.[7] Consequently, water emissions from the Wellman-Lord pre-scrubber stream should thus be limited to intermittent discharges from the ash pond.

The aqueous stream from a sodium throwaway system will contain about five percent dissolved solids. Discharge rates and average stream compositions for the cases considered in this study are given in Table 6.2-8. The major differences between the dissolved solid compositions of the eastern and western coal applications is due to the fact that 20 percent oxidation was assumed to occur in the western coal cases whereas 10 percent was assumed to occur for the eastern cases and 15 percent for the 2.3% S coal cases.

In sodium throwaway systems, the absorbed SO_2 reacts to form Na_2SO_3 and Na_2SO_4 which are removed from the system as dissolved solids in an aqueous waste. Consequently, the amount of aqueous emissions is directly related to both the SO_2 control level and the coal's sulfur concentration. The amount of aqueous wastes from high sulfur eastern coal applications will be approximately 4.7 times those from western coal-fired applications.

Some of the sodium throwaway systems in use today for controlling SO_2 emissions from industrial boilers use a waste process stream that contains sodium as a feed to the scrubber. Sodium scrubbers applied to boilers in some paper mills and textile plants are examples. The relatively small aqueous process waste stream from the FGD system is then recombined with the industrial process waste streams and discharged to a centralized water treating plant or a municipal sewer.

Common water treating practice for sodium throwaway systems in use today is to discharge their wastes to an evaporation pond or to an existing centralized water treating plant. Of the 102 sodium scrubbing systems in use today, about 80 use evaporation ponds and 10 use centralized water treating for disposal of their FGD wastes. The remaining systems use varied approaches ranging from discharge to city sewers to deep mine injection.[8]

TABLE 6.2-8. WATER POLLUTION IMPACTS FOR THE SODIUM THROWAWAY SYSTEM*

Boiler size and type	Control level	3.5% S eastern coal		2.3% S coal		0.6% S western coal	
		ℓ/sec	(gpm)	ℓ/sec	(gpm)	ℓ/sec	(gpm)
8.8 MW_t(30 X 10 Btu/hr) Underfeed stoker	90	0.85	13.4	-	-	0.17	2.7
	85	0.79	12.5	-	-	0.16	2.6
	75	0.71	11.2	-	-	0.15	2.4
	56	0.52	8.4	-	-	-	-
22 MW_t(75 X 10 Btu/hr) Chaingrate Stoker	90	2.18	34.6	1.28	20.3	0.46	7.3
	75	1.77	28.1	-	-	0.37	5.9
	56	1.26	19.9	-	-	-	-
44 MW_t(150 X 10 Btu/hr) Spreader Stoker	90	4.38	69.4	-	-	0.92	14.6
	75	3.53	55.9	-	-	0.75	11.8
	56	2.45	38.8	-	-	-	-
58.6 Mw_t(200 X 10 Btu/hr) Pulverized coal	90	5.80	92.0	-	-	1.23	19.5
	85	5.26	83.4	-	-	1.11	17.6
	75	4.71	74.7	-	-	0.99	15.7
	56	3.34	52.9	-	-	-	-
118 MW_t(400 X 10 Btu/hr) Pulverized coal	90	11.7	185	6.82	108	2.46	38.9
Avg. Dissolved Solid Compositions		Na_2SO_3= 77 percent Na_2SO_4= 9 percent NA_2CO_3= 14 percent		73 percent 13 percent 14 percent		69 percent 17 percent 14 percent	
Avg. TDS Concentration (wt. %)		5		5		5	

*Based on material balance calculations provided in Appendix A.

For purposes of this evaluation, onsite treatment of sodium system aqueous wastes using a basic water treating scheme of sulfite oxidation and pH neutralization was selected as the treatment method for evaluation. Although evaporation ponds are currently used in the majority of sodium system applications, their use is limited to certain geographic areas of the county where the annual evaporation rate exceeds the annual rainfall. The water treatment systems selected for this evaluation will result in a sodium sulfate stream which must be disposed.

6.2.3 Solid Waste

The major solid waste impacts from the five candidate processes result from the sludges produced in the limestone and double-alkali processes and the dry solid produced in the spray drying process. A solid purge stream of Na_2SO_4 is produced in the Wellman-Lord process, but the stream is relatively small and should not constitute a major solid waste impact, especially for the size applications under consideration in this study.

Both the limestone and double alkali sludges are composed primarily of calcium sulfite and sulfate salts. Significant amounts of fly ash may also be present, depending on the method of upstream particulate control in use. The sludges are relatively inert and with proper site selection and proper disposal procedures, can be disposed of in an environmentally acceptable manner. The disposal methods currently in use are lined and unlined ponding and landfilling of treated and untreated materials. Potential adverse impacts of sludge disposal lie in the areas of disposal acreage requirements, water contamination through leaching and percolation of soluble components of the solid waste into the groundwater system, and land use impacts due to poor structural properties. Treatment techniques to minimize adverse impacts may involve dewatering, addition of alkaline ash, and/or application of commercial stabilization technology. These techniques are used to decrease the sludge volume, decrease its permeability, and improve its structural properties as discussed in the following sections. In some instances, forced oxidation can also be considered a sludge beneficiation technique as it results in a more easily dewatered material.

The dry solid waste product from the spray drying process will consist primarily of calcium or sodium salts, depending upon the type of alkali used as the SO_2 sorbent. Significant amounts of fly ash will also be present since the solids collection device associated with the spray drier, probably a baghouse, will remove the particulates generated from the coal combustion process along with the spray drying solid wastes. Upstream particulate removal is not practical for this process since the spray dryer's performance is not adversely affected by the presence of fly ash[10] and dual particulate removal units would be unattractive from both an energy and economic view. The alkali, coal, and ash compositions have impacts on waste disposal alternatives which include disposal in both lined and unlined landfills and the potential for producing a low quality cement.[11]

Volume of Sludge Production

As with the sodium throwaway system, all of the SO_2 absorbed from the flue gas by a limestone system must leave the process in a waste stream, in this case as a waste sludge. Consequently, the amount of sludge produced by a limestone system is proportional to the sulfur content of the coal and the SO_2 removal level. Table 6.2-9 presents the results of the limestone process material balance calculations and shows the variation in sludge production with coal sulfur content and SO_2 removal. Sludge production rates presented in this section are on an ash-free basis, since upstream particulate removal was assumed for this study.

The volume of sludge produced is also important as the sludge volume will determine the size of the holding pond or landfill area. Table 6.2-10 presents the results of calculations to estimate the sludge volumes produced by a limestone process for the standard sized boilers. The difference in sludge densities for the two coals is due to the higher oxidation for the western coal cases. Results are presented in units of cc/sec, lb/hr, and acre-feet/30 years. The last category, acre-feet/30 years gives an indication of the total volume of sludge to be handled over the life of the plant assuming a 30 year life and an onstream factor of 60 percent. Figure 6.2-5 illustrates

TABLE 6.2-9. SOLID WASTE IMPACT FOR THE LIMESTONE FGD PROCESS
(Ash-Free Basis)

Boiler size and type	Percent removal	3.5% S eastern coal				0.6% S western coal			
		g/s	(lb/hr)	ℓ/min	(gal/min)	g/s	(lb/hr)	ℓ/min	(gal/min)
8.8 MW_t (30×10^6 Btu/hr)	90	89.8	(712)	2.7	(0.7)	20.9	(166)	0.61	(0.16)
Underfeed Stoker	85	84.7	(672)	2.7	(0.7)	19.4	(154)	0.57	(0.15)
	75	74.9	(594)	2.3	(0.6)	17.4	(138)	0.53	(0.14)
22 MW_t (75×10^6 Btu/hr)	90	226.5	(1796)	6.8	(1.8)	52.2	(414)	1.5	(0.4)
Chaingrate Stoker	75	190.2	(1508)	5.7	(1.5)	43.6	(346)	1.1	(0.3)
44 MW_t (150×10^6 Btu/hr)	90	457.0	(3624)	13.6	(3.6)	104.7	(830)	3.0	(0.8)
Spreader Stoker	75	380.6	(3018)	11.4	(3.0)	87.3	(692)	2.7	(0.7)
58.6 MW_t (200×10^6 Btu/hr)	90	606.8	(4812)	18.2	(4.8)	139.5	(1106)	4.2	(1.1)
Pulverized Coal	85	576.1	(4568)	17.4	(4.6)	131.9	(1046)	3.8	(1.0)
	75	507.0	(4020)	15.2	(4.0)	116.0	(920)	3.4	(0.9)

TABLE 6.2-10. SOLID WASTE VOLUMES FOR THE LIMESTONE FGD PROCESS
(Ash-Free Basis)

Boiler size and type	Percent removal	3.5% S eastern coal* Sludge volume cc/sec	3.5% S eastern coal* Sludge volume (ft^3/hr)	3.5% S eastern coal* Sludge volume (acre-ft/ 30 yrs)	0.6% S western coal† Sludge volume cc/sec	0.6% S western coal† Sludge volume (ft^3/hr)	0.6% S western coal† Sludge volume (acre-ft/ 30 yrs)
8.8 MW_t(30×10^6 Btu/hr)	90	69.1	8.8	31.8	13.8	1.8	6.6
Underfeed Stoker	85	65.2	8.3	30.0	12.8	1.6	5.8
	75	57.6	7.3	26.4	11.5	1.5	5.4
22 MW_t(75×10^6 Btu/hr)	90	174.2	22.1	80.0	34.6	4.4	16.0
Chaingrate Stoker	75	146.3	18.6	67.4	28.9	3.7	13.4
44 MW_t(150×10^6 Btu/hr)	90	351.5	44.6	161.4	69.3	8.8	32.0
Spreader Stoker	75	292.8	37.2	134.6	57.8	7.4	26.8
58.6 MW_t(200×10^6 Btu/hr)	90	466.8	59.3	214.6	92.4	11.8	42.8
Pulverized coal	85	443.2	56.3	203.8	87.4	11.1	40.2
	75	390.0	49.5	179.2	76.8	9.8	35.4

*Eastern sludge bulk density = 1.30 g/cc (81.2 lb/ft^3)
†Western sludge bulk density = 1.51 g/cc (94.0 lb/ft^3)
} These are average values from Table 6.2-12. which follows.

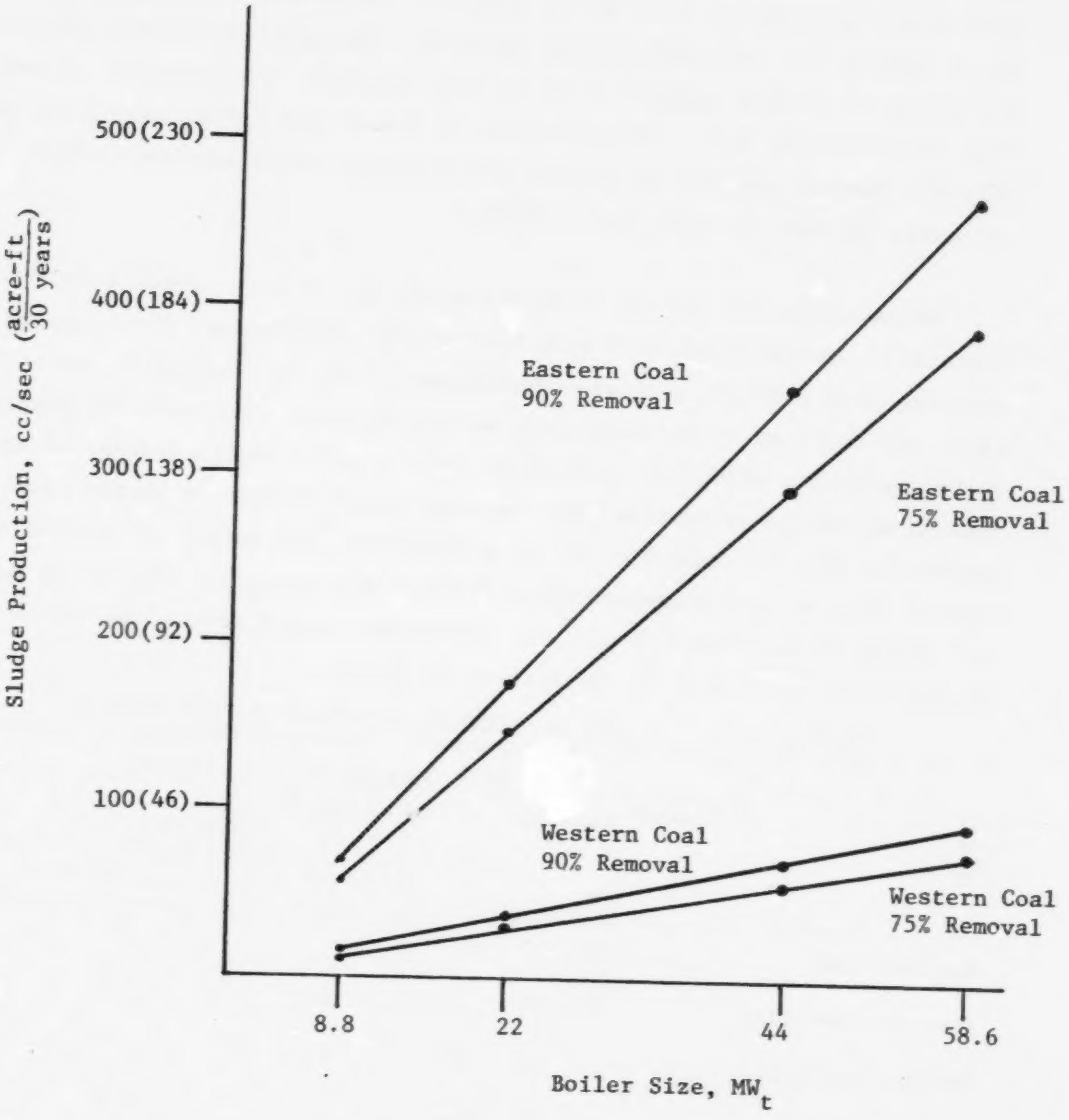

Figure 6.2-5. Sludge Production Rates for the Limestone FGD Process

the results of these calculations graphically and shows the variation in sludge production with coal sulfur content, boiler size, and level of removal. As one would expect, sludge production increases with all of these factors. The quantity of sludge produced from the double alkali process will also vary with sulfur removal and fuel sulfur content. Table 6.2-11 presents the solid waste impacts for the double-alkali process. The major difference between the amount of sludge produced from the double-alkali and limestone systems is that the limestone system stoichiometry is based upon 1.2 moles sorbent per mole SO_2 removed whereas the double-alkali system stoichiometry is based upon 1.0 moles sorbent per mole SO_2 removed.

Volume reduction can be accomplished by one or more sludge dewatering practices. Methods that have been tested are: thickening, discharge to settling pond (with or without underdrainage), vacuum filtration, centrifugation, and addition of dry materials such as fly ash. Although the degree of volume reduction which can be attained with a given sludge depends on several individual sludge properties, the relative effectiveness of various dewatering methods is seen by comparing the bulk densities (wet basis) of treated sludges. Several sets of data representing various sludge types and dewatering methods are presented in Table 6.2-12. The commercial identities of the various methods were not listed in the referenced report.

TABLE 6.2-12. EFFECT OF CHEMICAL FIXATION ON BULK DENSITY[12]

	Bulk density (lbs/ft)						
	Untreated	Treated*					
		1	2	3	4	5	Average of untreated and beneficial treatments for limestone
Eastern lime	52	100	77	101	-	-	-
Eastern limestone	63	108	89	83	-	63	81.2
Eastern dual alkali	52	96	91	99	-	53	-
Western limestone	89	109	80	111	81	57	94.0
Western dual alkali	47	97	82	83	-	68	-

*Each column represents one of five different commercial fixation treatments tested.

TABLE 6.2-11. SOLID WASTE IMPACT FOR THE DOUBLE ALKALI PROCESS (Ash free Basis)

Boiler Size and type	Percent remova	3.5% S eastern coal				0.6% S western coal				2.3% S coal			
		g/s	(lb/hr)	ℓ/min	(gal/min)	g/s	(lb/hr)	ℓ/min	(gal/min)	g/s	(lb/hr)	ℓ/min	(gal/min)
8.8 MW_t(30 X 10^6 Btu/hr) Underfeed Stoker	90	78.7	(624)	2.3	(0.6)	15.4	(122)	0.4	(0.1)	-	-	-	-
22 MW_t(75 X 10^6 Btu/hr) Chaingrate Stoker	90	200.0	(1584)	5.7	(1.5)	38.1	(302)	1.1	(0.3)	112.0	(888)	3.0	(0.8)
58.6 MW_t(200 X 10^6 Btu/hr) Pulverized Coal	90	506.0	(4012)	14.4	(3.8)	101.6	(806)	2.7	(0.7)	-	-	-	-
118 MW_t(400 X 10^6 Btu/hr) Pulverized Coal	90	1065.6	(8450)	30.7	(8.1)	203.5	(1614)	5.3	(1.4)	596.8	(4732)	17.1	(4.5)

One of the sludge parameters which determine ease of dewaterability is the relative sulfate/sulfite contents in the solids. High sulfate sludges (*i.e.*, with sulfate present as gypsum) tend to dewater much better than sludges with high sulfite levels. This in turn reduces the volume of material for disposal and makes a material which may be suitable for landfill without need of additives.[13] The EPA currently has studies underway at Shawnee to investigate forced oxidation as a means of achieving a waste with better settling and dewatering characteristics. Additional activities planned in this area include a full scale demonstration of the technology at TVA's Widows Creek plant.

All of the SO_2 absorbed in a spray dryer must also exit the process as a waste stream, in this case as a solid salt. Table 6.2-13 presents the quantity of solid wastes produced by the spray drying system as a function of SO_2 removal, coal type, and process size. Solid waste quantities from this process are, however, a combination of spray dryer solids and fly ash generated from coal combinations. It is interesting to note that for the cases considered in this evaluation, the majority of solid wastes from this process resulted from fly ash and not from the removal of SO_2. Table 6.2-14 presents a breakdown of the origin of solid waste material for each of the spray drier cases.

Results of pilot plant testing reported by Basin Electric were that the spray drying product produced from the coals they tested handled as well as fly ash and would not require special handling equipment other than the conventional dry handling equipment used for fly ash.[14] Disposal methods planned for the systems they have under construction are disposal in depleted mines and landfill after mixing with conventional scrubber sludge.[15] The two spray drying systems under construction for industrial boiler applications both plan to truck the waste solids to an offsite landfill area.[16,17]

TABLE 6.2-13. SOLID WASTES FROM SPRAY DRYING

(Total Fly Ash + Alkali Salts)

MW_t	Boiler size (10^6 Btu/hr)	% S Coal	SO_2 Removal	Sorbent	Sorbent Stoichiometry	Solids		Volume[a]		
						g/sec	(lb/hr)	cc/sec	(ft^3/hr)	acre-ft/15 years
44	(150)	0.6	90	Sodium	1.1	132	(1047)	111	(14.1)	25.5
44	(150)	0.6	75	Sodium	0.8	114	(904)	96	(12.2)	22.1
44	(150)	0.6	50	Sodium	0.5	99	(785)	83	(10.6)	19.2
58.6	(200)	0.6	75	Sodium	0.8	174	(1378)	146	(18.5)	33.5
17.6	(60)	0.6	75	Sodium	0.8	29	(227)	24	(3.1)	5.6
44	(150)	0.6	90	Lime	2.0	134	(1066)	113	(14.3)	25.9
44	(150)	0.6	75	Lime	1.2	114	(905)	96	(12.2)	22.0
44	(150)	0.6	50	Lime	0.65	96	(764)	81	(10.3)	18.6
118	(400)	0.6	70	Lime	1.2	344	(2725)	289	(36.7)	66.5
118	(400)	2.3	70	Lime	1.2	729	(5782)	613	(77.9)	141.0
22	(75)	2.3	70	Lime	1.2	85	(675)	71	(9.1)	16.5

TABLE 6.2-14. SOLID WASTE BY ORIGIN

						Amount of solid waste					
						Fly ash			Desulfurization products		
MW_t	(10^6 Btu/hr)	% S Coal	SO_2 Removal	Type sorbent	Sorbent Stoichiometry	g/s	(lb/hr)	Percent of total	g/s	(lb/hr)	Percent of total
44	(150)	0.6	90	Sodium	1.1	69	(546)	52	63	(501)	48
44	(150)	0.6	75	Sodium	0.8	69	(546)	60	43	(341)	40
44	(150)	0.6	50	Sodium	0.5	69	(546)	70	30	(239)	30
58.6	(200)	0.6	75	Sodium	0.8	113	(896)	65	61	(482)	35
17.6	(60)	0.6	75	Sodium	0.8	11	(84)	37	18	(143)	
44	(150)	0.6	90	Lime	2.0	69	(546)	51	66	(520)	49
44	(150)	0.6	75	Lime	1.2	69	(546)	60	45	(359)	40
44	(150)	0.6	50	Lime	0.65	69	(546)	71	27	(218)	29
118	(400)	0.6	70	Lime	1.2	226	(1791)	66	118	(934)	34
118	(400)	2.3	70	Lime	1.2	401	(3182)	55	328	(2600)	45
22	(75)	2.3	70	Lime	1.2	24	(187)	27	61	(488)	

Secondary Water Pollution

Secondary water pollution impacts from the disposal of lime/limestone, double alkali, and spray drying process solid wastes can occur primarily by landfill runoff, and/or leaching. Currently, there is very little data existing on the secondary water effects from spray drying wastes due to the early development stage of the technology. However, because of the similarity of the chemical composition of the products from a lime spray drying process and the sludge produced from a lime/limestone process, it has been suggested that their disposal problems may also be similar. Consequently, the following discussion on water impacts from calcium based sludge producing processes should generally hold for calcium based spray drying operations. However, the waste product from a sodium based spray drying system may require special handling because of its increased solubility. Acceptable disposal practices for spray drying products is an area currently undergoing further investigation.

Effects on surface waters by either runoff or overflow from FGD waste disposal operations are minimized because the runoff/overflow is mixed with the natural surface water and diluted by incident rainfall. The magnitude of this impact will depend on site-specific hydrological conditions. This impact, however, is not receiving as much emphasis as groundwater contamination because of decreasing probability of incidences occurring. This is due to zero discharge measures such as recycle of liquor from dewatering and ponding operations or treatment of concentrated bleed streams. Only in the event of unusually heavy rainfall or flooding conditions would this become a significant problem. Present regulations would prevent contaminated runoff from any storm event less than the 10 year-24 hour storm to be discharged without receiving treatment at least for solids and pH control.

In rare cases where upset conditions or flooding occurs and water needs to be purged from the holding pond, the purge streams will be saturated with dissolved calcium salts and may contain high concentrations of chloride, sodium, and other ions. Currently available water treating technology as discussed in Section 6.2.2 would be applicable to treating any purge liquors from sludge holding ponds.

Another potential impact of limestone and double alkali sludges is that the surrounding water quality can be impacted through leaching and percolation of soluble components of the solid waste into the groundwater system. Some of the liquor disposed of in unlined sludge ponds and landfills will either evaporate, percolate down through the sludge and subsoil and potentially to an underlying aquifer, or overflow the pond. Additional rainfall inflow will add to the total liquor volume of the contents unless covered. In unlined pond/landfills, as the liquor moves through the sludge solids, the dissolved solids components of the liquor and species leached from the solids are carried into the underlying subsoil. The leachate rate of any given pollutant is dependent on many factors such as: its solubility in the liquor percolating through the material, the permeability of the waste, the areas of the landfill or pond, the permeability of the soil, rainfall, and the hydraulic head.

The permeability of ponded or landfilled material is a measure of the rate at which leachate passes through a disposal site. Lower permeabilities result in lower leaching rates. This parameter is generally a function of the particle or grain size, void ratio, shape and arrangement of pores, the degree of saturation, fluid viscosity and temperature. Permeability directly affects the volume of leachate and, in case of high permeability, can result in physical abrasion of the fill itself.

Permeability coefficients for various FGD sludges range from 10^{-4} to 10^{-5} cm/sec, as shown in Table 6.2-15. According to these investigators, permeability is generally lower for sulfite sludges although close control of gypsum formation in a dual alkali system can yield a low permeability as shown by the ADL pilot plant sulfate dual alkali sludge.[18] Table 6.2-16 gives a comparison of permeabilities of settled or drained sludges with compacted sludges. In general, permeability was shown to decrease with compacted

TABLE 6.2-15. COEFFICIENTS OF PERMEABILITY FOR FGD SLUDGES[22]

Sludge	% Solids	Coefficient of permeability (cm/sec)
ADL Pilot Plant (sulfate dual alkali)	72.0	2.1×10^{-5}
Shawnee TVA Plant (sulfite-rich lime)	45.0 50.0	2.0×10^{-4} 5.0×10^{-5}
Shawnee TVA Plant (sulfite-rich limestone)	55.0 67.0	2.5×10^{-4} 6.0×10^{-5}
Duquesne Phillips Plant (sulfite-rich)	64.0 73.0	1.5×10^{-4} 6.0×10^{-5}
SCE Mohave Plant (100% sulfate)	73.0 85.0	6.0×10^{-4} 1.5×10^{-4}

TABLE 6.2-16. PERMEABILITIES OF FGD SLUDGES [23,24]

		Settled or drained		Compacted	
Location	Process	Void ratio	Coef. of perm. (cm/sec)	Void ratio	Coef. of perm. (cm/sec)
Eastern	Limestone				
	Sample 1	1.53	1×10^{-4}	1.24	8×10^{-5}
	Sample 2	2.07	3×10^{-5}	1.56	1×10^{-5}
Eastern	Lime				
	Sample 1	1.83	2×10^{-4}	1.68	5×10^{-5}
	Sample 2	1.65	6×10^{-5}	1.42	1×10^{-5}
	Sample 3	1.25	1×10^{-4}	0.97	7×10^{-5}
Western	Limestone				
	Sample 1	0.96	3×10^{-5}	0.63	1×10^{-5}
	Sample 2	1.20	2×10^{-5}	1.20	1×10^{-5}
	Sample 3	0.75	8×10^{-4}	0.50	9×10^{-5}
Eastern	Dual Alkali				
	Sample 1	5.11	8×10^{-5}	4.17	3×10^{-5}
	Sample 2	2.19	2×10^{-4}	1.95	8×10^{-5}
Western	Dual Alkali	2.77	1×10^{-3}	2.61	1×10^{-4}

sludge.[19,20,21] Permeability can also be reduced by addition of fly ash, lime, and/or commercial fixative as shown in Tables 6.2-17 and 6.2-18.

The composition of the leachate is a function of several factors including chemical composition of the sludge, pH, solubility of the individual species present, and age of the disposal site. The nature of the leachate evolving from untreated sludge can be judged by analysis of liquors associated with scrubber samples (especially scrubber sludge discharge liquors), laboratory-generated leachate results, or actual field leachate samples. Since fly ash will be collected and/or disposed of and mixed with the scrubber sludge in many cases, it is likely that many of the potentially leachable elements originate with the ash.

Although the compositions of sludge and sludge-ash leachates can vary significantly, an overview of the nature of sludge liquors can be seen in Table 6.2-19 and 6.2-20. These data represent liquors from many types of FGD systems, coal sources and geographic regions. Chloride and sulfate values are high and may represent the most significant problem in disposing of FGD sludges. Since these liquors have to be diluted by rainwater before release to the environment, the value of detailed comparisons is limited. TDS levels ranged as high as 95,000 mg/ℓ . The pH's reported were mostly all neutral or alkaline. Many elements varied by as many as three orders of magnitude from sludge to sludge, including As, Be, Cr, Cu, Pb, Mn, Mo, Ni, Se, Na, and Zn. From the scattering of data presented, it was not possible to identify any correlatable parameters from adherent sludge liquor analyses. Some indication was evident that trace element levels may be higher in eastern sludge.

Little documentation is available to determine whether or not leachate from operating ash disposal sites is a potential source of pollution. However, this is not to be interpreted that a pollution problem does not exist. The composition of the fly ash, the concentration of soluble constituents, the liquor pH, the trace metal solubility, soil attenuation, and/or the lateral water movement in underlying aquifers may be such that the problem area has

TABLE 6.2-17. COEFFICIENTS OF PERMEABILITY FOR FGD SLUDGES[25] TREATED WITH FLY ASH AND/OR CEMENT

Sludge	% Solids	Coef. of perm. (cm/sec)
Sludge + 5% cement	66	7.6×10^{-6}
Sludge + 65% fly ash	79	7.2×10^{-6}
Sludge + 35% fly ash + 5% cement	75	4.0×10^{-6}
Sludge + 39% fly ash + 5% lime	75	5.6×10^{-6}

TABLE 6.2-18. EFFECT OF SLUDGE TREATMENT ON PERMEABILITY[26]

Location	Process	Compacted untreated	Coefficient of permeability (cm/sec)				
			Process "A"	Process "B"	Process "E"	Process "F"	Process "G"
Eastern	Lime	1×10^{-5}	2×10^{-6}	2×10^{-4}	8×10^{-4}	-	-
Eastern	Limestone	8×10^{-5}	-	1×10^{-5}	3×10^{-6}	-	5×10^{-5}
Eastern	Dual alkali	3×10^{-5}	1×10^{-7}	5×10^{-5}	5×10^{-11}	-	1×10^{-4}
Western	Limestone	1×10^{-5}	4×10^{-7}	4×10^{-5}	4×10^{-8}	5×10^{-6}	1×10^{-4}
Western	Dual alkali	7×10^{-5}	9×10^{-7}	7×10^{-5}	$- \times 10^{-7}$	-	4×10^{-5}

TABLE 6.2-19. EQUILIBRIUM CONCENTRATIONS OF TRACE ELEMENTS IN FGD SLUDGE LEACHATE (in ppm)[27]

Coal source: sulfur content, % scrubber absorbent: ash collection	Station 1 sludge Western low lime ESP upstream	Station 4 sludge Eastern 4 limestone ESP upstream	Station 5 sludge Eastern 5 limestone with sludge	Station 1 lime - -
pH	8.5	9.7	8.4	12.6
Antimony	0.014	0.013	0.035	0.016
Arsenic	<0.002	<0.002	0.03	<0.002
Barium	2	<0.3	<0.3	<0.3
Beryllium	0.002	0.001	0.002	0.001
Boron	2.6	6.3	0.96	0.22
Cadmium	0.0005	<0.001	0.002	<0.001
Chromium	0.001	0.011	<0.001	0.004
Copper	0.031	0.045	<0.005	0.013
Fluorine	31.5	8.7	7.6	1.2
Germanium	<0.01	<0.01	0.02	<0.01
Lead	0.0056	0.0033	0.0061	0.0027
Manganese	<0.002	<0.002	<0.002	<0.002
Mercury	0.0005	0.001	0.0008	0.002
Molybdenum	0.063	0.061	0.075	0.011
Nickel	<0.05	<0.05	<0.05	<0.05
Selenium	0.045	0.0096	0.016	0.0005
Vanadium	<0.1	<0.1	<0.1	<0.1
Zinc	0.005	0.052	<0.005	0.11

Source: Holand et al., 1975

Note: Underscored values indicate in excess of National Interim Primary Drinking Water Standards (NIPDWS).

TABLE 6.2-20. LEVELS OF CHEMICAL SPECIES IN FGD SLUDGE LIQUORS AND ELUTRIATES[28]

Species	Eastern coals			Western coals		
	Range in liquor (ppm)	Median (ppm)	Total no. of observations	Range in liquor (ppm)	Median (ppm)	Total no. of observations
Antimony	0.46-1.6	1.2	4	0.09-0.22	0.16	2
Arsenic	<0.004-1.8	0.020	15	<0.004-0.2	0.009	7
Beryllium	<0.0005-0.05	0.014	16	0.0006-0.14	0.013	7
Boron	41	41	1	8.0	8.0	1
Cadmium	0.004-0.1	0.023	11	0.011-0.044	0.032	7
Calcium	470-2,600	700	15	240-(∿45,000)*	720	6
Chromium	0.001-0.5	0.020	15	0.024-0.4	0.08	7
Cobalt	<0.002-0.1	0.35	3	0.1-0.17	0.14	2
Copper	0.002-0.4	0.015	15	0.002-0.6	0.20	7
Iron	0.02-0.1	0.026	5	0.[illegible]2-8.1	4.3	2
Lead	0.002-0.55	0.12	15	0.0014-0.37	0.016	7
Manganese	<0.01-9.0	0.17	8	0.007-2.5	0.74	6
Mercury	0.0009-0.07	0.001	10	<0.01-0.07	0.01	7
Molybdenum	5.3	5.3	1	0.91	0.91	1
Nickel	0.03-0.91	0.13	11	0.005-1.5	0.09	6
Selenium	<0.005-2.7	0.11	14	<0.001-2.2	0.14	7
Sodium	36-20,000†	118	6	1,650-(∿9,000)†	-	2
Zinc	0.01-27	0.046	15	0.028-0.88	0.18	7
Chloride	470-5,000	2,300	9	1,700-43,000†	-	2
Fluoride	1.4-70	3.2	9	0.7-3.0	1.5	3
Sulfate	720-30,000†	2,100	13	2,100-18,500†	3,700	7
TDS	2,500-70,000†	7,000	-	5,000-95,000*	12,000	3
pH	7.1-12.8	-	-	2.8-10.2	-	

*Levels of soluble chloride components in sludges are dependent upon the chloride-to-sulfur ratio in the coal. The highest levels shown are single measurements for a western limestone scrubbing system operating in a closed-loop using cooling tower blowdown for process makeup water.

†Levels of soluble sodium salts in dual alkali sludge (filter cake) depend strongly on the degree of cake wash. The highest levels shown reflect single measurements on an unwashed dual alkali filter cake.

Source: Lunt, et al. 1977

not been identified. Information on the chemical properties of sulfur oxide sludge indicates a potential pollution problem and a need for careful site selection, and in some cases installation of a liner or drainage system. Results of field studies now in progress have not indicated that serious problems exist with respect to FGD sludge leachate, particularly so with regard to trace elements. It can be assumed that FGD sludge landfills, at least at this time, involve placement of chemically treated wastes. In the future, however, landfilling of gypsum-type sludges generated under forced oxidation conditions might be feasible. Chemically fixed sludges have been found to generate leachates lower in TDS over short terms, but long term behavior is still a matter of concern.

Wellman-Lord Process

Table 6.2-21 presents the results of the calculations to estimate the quantity of solid waste materials produced in a Wellman-Lord process applied to the standard size boilers. As can be seen, the quantity of waste material is quite small varying from about 1 to 20 g/sec (9 to 180 lb/hr) depending upon the boiler size and the coal sulfur content. This material is essentially pure Na_2SO_4 and Na_2SO_3 for which a market may exist in the wood pulping and glass making industries. Consequently, due to the small waste stream size and its potential marketability, no significant solid waste impacts are foreseen for the boiler sizes in question. If this material can not be sold it can be disposed of in a lined and covered landfill area.

TABLE 6.2-21. SOLID WASTE PRODUCTION FOR THE WELLMAN-LORD PROCESS

Boiler size and type	Percent removal	Waste production 3.5% S easterm coal g/s	 (lb/hr)	 0.6% S western coal g/s	 (lb/hr)
8.8 MW_t (30×10^6 Btu/hr) Underfeed Stoker	90	4.4	(34.2)	1.2	(9.5)
22 MW_t (75×10^6 Btu/hr) Chaingrate Stoker	90	20.4	(162)	3.0	(23.7)
58.6 MW_t (200×10^6 Btu/hr) Pulverized Coal	90	22.6	(179)	7.9	(62.4)

6.2.4 Environmental Impact on Modified Facilities

The environmental impact of applying FGD controls to modified or reconstructed boilers will essentially be the same as for newly constructed boilers. Consequently, previous discussions of air, liquid, and solid waste impacts are applicable to modified and reconstructed facilities.

6.3 IMPACT OF CONTROLS FOR OIL-FIRED BOILERS

The environmental impact of FGD systems on oil-fired boilers will be similar to that for coal-fired boilers. The major difference will be that the oil-fired boilers will generally burn a lower sulfur fuel than the 3.5 percent sulfur eastern coal which will result in lower air, liquid, and solid emissions. The standard residual oil-fired boiler burns a 3.0 percent sulfur oil that produces uncontrolled SO_2 emissions of 213.4 kg/hr (471.0 lb/hr). The standard coal-fired boilers of the same size produced uncontrolled emissions of 283.2 kg/hr (845.9 lb/hr) for the 3.5 percent sulfur eastern coal and 8.67 kg/hr (178.1 lb/hr) for the 0.6 percent sulfur western coal. Consequently, all air, liquid, and solid emissions for the standard residual oil-fired boiler will be approximately half those of the high sulfur eastern coal and 2 1/2 those of the western coal. Most of the treatment methods discussed in the previous sections are equally applicable to oil-fired FGD applications, however, sludge stabilization techniques using fly ash from coal will not be practical for oil-fired installations.

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SECTION 7
EMISSION SOURCE TEST DATA

7.1 Introduction

The objective of this chapter is to present continuous monitoring data for FGD systems applied to industrial boilers that illustrate system performances during long term operations. These data would also show variations in SO_2 removal for 8 hour, 24 hour, and 30 day emission averaging times.

There are several FGD systems currently being used in the United States to treat flue gases from industrial boiler; however, little continuous monitoring data were found to meet the objectives of this chapter. Discrete data sets from three EPA sponsored test programs are available which illustrate FGD system performance capabilities for industrial boiler applications.[1,2,3] Data from these FGD systems, listed in Table 7.1-1, are presented and discussed in this section.

TABLE 7.1-1. SOURCE MONITORING DATA FROM INDUSTRIAL BOILER FGD SYSTEMS

Boiler location	FGD type	Data status
Rickenbacker A.F.B. Columbus, Ohio	Lime/Limestone	Test data collected from March 1976 to May 1977.
Firestone Tire and Rubber Co.* Pottstown, Pennsylvania	Double Alkali	Test data collected from 9/27/77 to 10/8/77.
General Motors* Parma, Ohio	Double Alkali	Test data collected from 8/19/74 to 9/13/74, 2/17/75 to 3/14/75, and 4/19/76 to 5/14/76.

*These systems had the ability to burn either coal or oil. Data for the Firestone installation were collected while burning both fuel types.

The best examples of 30-day continuous monitoring data for industrial boiler FGD systems presently available are from a sodium throwaway system and a Wellman-Lord system presently operating on industrial boilers in Japan.[4] Dr. Jumpei Ando graciously provided Radian with these data from 71 MW_e and 135 MW_e installations, respectively, each firing 3 percent sulfur heavy oil. Plots of these data are shown in Section 7.4.

Continuous monitoring data from seven coal-fired utility FGD systems have recently been reported by EPA.[5,6,7] These systems include both regenerable and nonregenerable FGD processes and illustrated that SO_2 removals in excess of 85 percent are achievable over a 30 day operating period using technology that is available today. In addition to these data, EPA is currently obtaining continuous monitoring data from one additional lime/limestone systems. Also, data from performance tests conducted by an outside contractor on one sodium throwaway system are available. Table 7.1-2 presents a summary of the status of data gathering efforts for these nine utility FGD systems. Because of the

TABLE 7.1-2. CONTINUOUS MONITORING DATA FROM UTILITY BOILER FGD SYSTEMS

Utility	Plant site	FGD type	Data status
Louisville Gas & Electric	Cane Run Unit No. 4	Lime	89 days of continuous data
Pennsylvania Power Co.	Bruce Mansfield Unit No. 1	Lime	31 days of continuous data
Philadelphia Electric Co.	Eddystone Unit No. 1	Magnesium Oxide	5 days of continuous data
Northern Indiana Public Service Co.	D.H. Mitchell Unit No. 11	Wellman-Lord	56 days data reported. including 41 days continuous data
Columbus and Southern Ohio	Conesville Unit No. 5	Lime	34 intermittent days of data
Kansas City Power & Light	LaCygne Unit No. 1	Limestone	30 day monitoring test in progress
Kansas Power & Light	Lawrence Unit No. 4	Lime	22 days continuous data reported
Tennessee Valley Authority	Shawnee Test Facility	Limestone	42 days continuous data reported
Nevada Power Company	Reid Gardner Units No. 1 and 2	Sodium Throwaway	6 days data reported including 3 days continuous data

similiarity in flue gases from utility and industrial boilers, the utility data should be useful to indicate the performance levels that can be expected of industrial sized FGD systems.

7.2 Emission Source Data for Coal-Fired Boilers

In this section, brief descriptions of the test facilities will be presented, the test results will be discussed, and test procedures will be described. More detailed descriptions and discussions of the FGD processes being considered were presented in Section 2.

Data for a lime scrubbing process were collected at the R-C/Bahco scrubbing system installed at the Central Heat Plant at Rickenbacker Air Force Base near Columbus, Ohio. The heat plant houses eight coal-fired hot water generators with a total fuel burning capacity of approximately 330 x 10^6 Btu/hr. These stoker-fired generators burn 2.5 to 3.5% sulfur Ohio coal with an average heating value of 11,300 Btu/lb.[8]

The R-C/Bahco system was designed to treat up to 108,000 acfm of flue gas generated at the peak winter load of approximately 200 x 10^6 Btu/hr. The system, which must operate over the relatively narrow range of gas flow of 35,000 to 50,000 scfm, has an essentially unlimited turndown capability for handling flue gas by mixing air with the flue gas at low boiler loads. This allows the system to handle seasonal load variations from 20 to 200 x 10^6 Btu/hr, SO_2 concentrations from 200 to 2000 ppm and particulate loadings of up to 2 gr/scfd.[9]

Data for double-alkali systems were collected from an F.M.C. system located at Firestone Tire and Rubber Company's Pottstown, Pennsylvania facility. The test boiler for the FMC system is one of four comprising a steam plant which supplies process steam and heating steam for the facility. The boiler is one of three which operates at a fairly constant rate of 45,400 kg/hr (100,000 lb/hr) of steam. Process steam demand is relatively steady, since the plant operates 24 hours per day, seven days per week. Fluctuations in heating load are satisfied by either boosting steam generation rates on these boilers or by operating the fourth boiler (No. 1). The steam generation rate of Boiler No. 1 varies from zero to approximately 22,700 kg/hr (50,000 lb/hr) of steam.[10]

The boiler was installed in 1958 and was originally designed as a coal-fired unit. It was converted to fire either coal or fuel oil in 1967. The two fuels are usually not burned simultaneously except when converting from oil to coal firing. The coal is ignited by continuing oil firing until a stable coal flame is obtained. Oil and coal can be fired simultaneously to maintain acceptable steam generation rates if coal with a low heat content is burned.[11]

The flue gases are treated by an air pollution control system consisting of multiclone units for particulate control and a pilot double alkali FGD unit designed by FMC. All of the flue gas passes through the multiclones. The stream then is split and two-thirds of the flue gas ducted to the stack. The other one-third is ducted to the pilot FGD system. The boiler has no NO_x controls.

Data for double-alkali systems were also collected from the General Motors system located at their Chevrolet Parma, Ohio plant. Scrubbers were started up in March 1974 on each of the four boilers in the steam plant. Two boilers have steam generation rates of 27,300 kg/hr (60,000 lb/hr) and two have rates of 45,500 kg/hr (100,000 lb/hr). All boilers are spreader stoker-fired, normally burning high sulfur (2-3% S) eastern coal and, on occasion, lower sulfur waste oil.

The flue gases are first treated by existing mechanical dust collectors on each boiler for primary particulate control. Then the gases enter the scrubbers which are three-tray columns since only modest reduction of particulates is required.

7.3 Data Presentation

Performance data for the R-C/Bahco FGD system is presented in three tables which represent three separate test periods in the overall program to evaluate the performance of the R-C/Bahco system. Table 7.3-1 presents the initial SO_2 removal data which were taken during startup and early operation of the system (Spring 1976). These initial removal efficiencies ranged from 87 to 99 percent and were reported to be affected primarily by the lime/SO_2 stoichiometry.

A series of lime screening tests were then carried out from December 1976 until February 1977. The objective of these tests was to identify the effects of major process variables upon SO_2 removal. Consequently, various independent process variables such as gas flow rate, pressure drop, liquid pumping rate, stoichiometry, and slurry concentration were varied in a series of 21 discrete tests to examine their effect on SO_2 removal. Table 7.3-2 presents a summary of this test data. These tests indicated that the only variable of significance affecting SO_2 removal at the 95 percent confidence level was lime/SO_2 stoichiometry. Conclusions from the lime testing were:

1) Virtually any desired SO_2 removal efficiency can be achieved in the R-C/Bahco scrubber, when using lime, simply by adjusting the lime/SO_2 stoichiometry.

2) Lime utilization approaching 100 percent is achieved at stoichiometric ratios's of up to about 0.9. At stoichiometric ratios up to 1.1, producing up to 99 percent SO_2 removal, lime utilization is above 90 percent.[14]

In the Spring of 1977, a second series of variable tests were run using limestone as the sorbent. This test series was modeled after the lime screening test and examined the effects of the same process variables. Table 7.3-3 presents a summary of the limestone screening test data. Conclusions from the

TABLE 7.3-1. SO_2 REMOVAL EFFICIENCY DATA[12]

1976 Date	Coal sulfur content	Coal firing rate (MM Btu/hr)	Inlet SO_2 concentration (ppm)	Outlet SO_2 concentration (ppm)	SO_2 removal efficiency (%)	SO_2 emission rate (lbs/MM Btu)	Lime utilization (%)
3/30	3.24	132.2	1,392	156	87.6	0.621	100.0
4/8	3.24	115.2	1,200	45	95.7	0.21	94.0
5/19	3.25	47.9	454	24	94.4	0.29	98.8
5/26	2.64	54.0	555	5	99.0	0.045	90.3
5/26	2.64	52.8	489	8	98.2	0.084	90.5
5/26	2.64	43.3	401	8	97.9	0.095	91.2
5/27	2.01	44.8	327	5	98.3	0.061	94.2
5/27	2.01	44.3	323	5	98.2	0.061	95.4

Note: Removal efficiencies were corrected for increased outlet gas volume due to water evaporation in the scrubber.

TABLE 7.3-2. LIME TEST DATA SUMMARY - RICKENBACKER AIR FORCE BASE - R-C/BAHCO LIME SCRUBBING SYSTEM[13]

Test No.	and date	Coal firing rate (lb/hr)	Average lime/SO_2 stoichiometry	Flue gas flow rate (10^3 acfm @ 120°F)	Average SO_2 (ppm) Inlet	Average SO_2 (ppm) Outlet	Overall SO_2 control efficiency (percent)	Average SO_2 emissions ng/J	Average SO_2 emissions (lb/10^6 Btu)
1	12/16/76	14,098	0.94	56.2	1,110	104	89.4	146.2	0.34
2R*	2/17/77	13,213	0.60	46.2	1,500	550	59.6	675.0	1.57
3	12/17/76	14,994	1.00	47.6	1,000	28	96.8	30.1	0.07
4	12/15/76	13,991	0.98	65.4	992	40	95.4	60.2	0.14
5R*	2/17/77	13,361	0.65	68.1	1,140	365	64.5	889.6	2.07
6	2/16/77	13,214	0.93	46.3	1,100	81	92.5	103.2	0.24
7R*	2/18/77	13,368	0.60	47.7	1,350	494	59.8	640.6	1.49
8	2/17/77	13,229	0.36	66.1	1,275	742	35.9	1298.4	3.02
9	12/15/76	12,712	0.97	68.6	910	58	92.7	103.2	0.24
10	12/18/76	13,229	0.99	55.7	1,075	41	95.6	60.2	0.14
11	12/18/76	14,627	0.98	57.9	1,150	70	93.3	94.6	0.22
12	12/19/76	15,522	0.99	70.2	842	40	94.6	64.5	0.15
13	12/19/76	10,789	0.81	67.6	940	174	78.8	386.9	0.90
14R*	2/13/77	14,983	0.54	45.9	1,200	546	54.5	610.5	1.42
15	12/13/76	16,639	1.06	49.3	1,500	25	98.1	25.8	0.06
16	2/13/77	10,448	0.66	70.2	1,095	340	65.8	834.0	1.94
17	2/15/77	15,878	0.92	64.5	1,000	115	87.5	180.6	0.42
18	12/20/76	20,313	1.08	50.4	1,155	14	98.7	12.9	0.03
19	2/15/77	12,539	0.67	45.2	1,045	348	63.4	451.4	1.05
20	2/10/77	8,954	0.60	53.4	950	370	57.8	451.4	1.80
21	2/14/77	10,430	0.85	54.9	1,075	172	82.3	318.1	0.74

*The "R" designation indicates a repeat test

TABLE 7.3-3. LIMESTONE TEST DATA SUMMARY - RICKENBACKER AIR FORCE BASE - R-C/BAHCO SCRUBBING SYSTEM [15]

Test No. and date		Coal firing rate (lb/hr)	Average Limestone/SO_2 stoichiometry	Flue gas flow rate (10^3 acfm @ 120°F)	Average SO_2 (ppm)		Overall SO_2 control efficiency (percent)	Average SO_2 emissions	
					Inlet	Outlet		ng/J	(lb/10^6 Btu)
	(1977 dates)								
38	5/19	4312	0.80	48.8	438	109	75.7	438.6	1.02
39	5/23	4928	1.53	36.0	425	30	88.8	77.4	0.18
40	5/24	4531	1.55	42.8	350	33	81.8	107.5	0.25
41	5/20	3955	1.42	52.2	290	31	92.2	146.2	0.34
42	5/23	3338	1.08	54.2	300	92	72.9	528.9	1.23
43	5/25	4158	1.41	39.5	360	23	94.6	81.7	0.19
44	5/25	4992	1.88	44.9	410	57	80.9	189.2	0.44
45	5/25	5252	1.30	60.2	320	35	89.0	146.2	0.34
46	5/25	3984	1.14	64.0	295	70	69.7	412.8	0.96
47	5/19	5006	0.86	51.2	390	90	86.5	326.8	0.76
48	5/1	3949	0.95	45.4	410	71	76.7	288.1	0.67
49	5/1	4765	1.19	54.2	375	90	84.7	335.4	0.78
50	5/1	3958	0.63	54.2	254	125	79.4	636.4	1.48
51	5/1	3960	0.96	37.0	483	102	90.2	322.5	0.75
52	5/1	3631	0.72	43.3	425	194	65.3	786.9	1.83
53	5/17	5408	1.19	57.5	350	71	84.4	266.6	0.62
54	5/17	4165	0.94	58.9	375	131	64.4	649.3	1.51
55	5/1	4158	1.38	32.6	600	107	84.8	296.7	0.69
56	5/1	4092	0.59	39.3	525	284	50.3	958.9	2.23
57	5/19	3393	0.94	465	375	69	81.7	335.4	0.78
58	5/19	3966	1.01	45.9	335	55	81.1	223.6	0.52

1) Limestone/SO_2 stoichiometry and slurry pumping rate are the significant variables contolling SO_2 removal efficiency.
2) A considerable excess of limestone is needed to absorbe SO_2 , especially at high SO_2 removal rates.
3) Limestone can be used to meet the requriements for SO_2 removal at RAFB (1.0 lb $SO_2/10^6$ Btu).[16]

Results of one week of testing at the Firestone test facility are presented in Table 7.3-4. Sulfur dioxide removal data from the Firestone facility indicated an average scrubber efficiency of about 97 percent. Controlled SO_2 emissions averaged 36.3 mg/J (0.08 pounds/10^6 Btu) which is less than either existing or proposed standards for utility boilers.

The results of the three one-month intensive testing programs at the GM Parma facility showed that the stringent level of SO_2 control (90% SO_2 removal) is obtainable by double alkali systems.[18] Some of the process variables studied include lime and soda ash stoichiometries, recycle pH, scrubber feed location, filter cake washing, and solids recycle.

Results of all three test programs indicate that high SO_2 removal efficiencies are achievable in coal-fired industrial boiler installations. Additional continuous monitoring data will be rquired, however, to evaluate industrial boiler FGD performance during long-term operations.

7.3.1 Test Methods

SO_2 data for the Rickenbacker tests were collected using the test method described in this section. SO_2 data for the Firestone tests were collected using a pulsed fluorescence analyzer.[19] Firestone test data reported in Table 7.3-4 are averages of the data collected during each test period. Continuous monitoring for the GM tests was reported to be an UV absorption analyzer with semi-continuous determination.

TABLE 7.3-4. COAL-FIRED EMISSION SOURCE DATA
FIRESTONE TIRE AND RUBBER - FMC DOUBLE-ALKALI FGD SYSTEM[17]

Test No.	Coal feed rate kg/hr	Coal heating value kg/kg	SO_2 inlet		SO_2 outlet		SO_2 control efficiency (percent)
			ng/J	lb/10^6 Btu	ng/g	(lb/10^6 Btu)	
200	3,629	29,263	1,009	2.35	25.4	0.06	97.5
201-1	3,629	28,872	1,284	2.99	39.0	0.09	96.9
201-2	3,629	29,997	1,295	3.01	35.5	0.08	97.2
201-3	3,175	29,419	1,028	2.39	31.8	0.07	96.9
201-4	3,629	29,878	942	2.19	49.7	0.12	94.7
Avg. Value	3,538	29,485	1,112	2.59	36.3	0.08	96.6

The following discussion describes the test method used at Rickenbacker A.F.B. to determine the flue gas SO_2 content. This method is only approximate and should be used only as a semiquantitative check on SO_2 concentrations. No temperature or pressure corrections have been incorporated, and the method should not be used below 100 ppm.[20]

Apparatus

1) 250 ml impinger with an open glass dip tube
2) A dry test meter
3) A source of vacuum
4) 25 ml pipette
5) Vacuum tubing
6) Hose clamp

Reagents

1) 3% Hydrogen Peroxide
2) 0.1N NaOH or 0.01N NaOH
3) Methyl/Orange-Xylene Cyanol indicator

Procedure:

Inlet Samples - (i.e., 500 + ppm SO_2) pipette 25 ml of 0.1N NaOH into the 250 ml impinger, add 50 ml of 3% hydrogen peroxide. Add approximately 25 ml of deionized water. Add several drops of Methyl/ Orange-Xylene Cyanol indicator.

Draw the gas sample through the impinger at 0.1 to 0.2 ft/min. Record the gas meter reading when the indicator turns from green to purple.

Outlet Samples - (100 to 600 ppm SO_2) substitute 0.01 normal NaOH for 0.1 normal NaOH in the above procedure. Follow the same procedure as above.

The following equation can be used to calculate the SO concentration:

$$SO_2 \text{ ppm} = \frac{10{,}000 \text{ x (NaOH Normality)}}{\text{Meter Volume ft}^3}$$

Note: Add the indicator within 15 minutes of running the test. If the indicator is added at an earlier time, it may be destroyed by the hydrogen peroxide in the impinger.

7.4 Emission Source Test Data for Oil-Fired Boilers

Three sets of data (including two sets of one month continuous testing) were found to illustrate SO_2 emissions from an oil-fired boiler. The Firestone facility previously discussed had provisions for firing both coal and oil. The FMC double alkali test system was operated under both oil- and coal-fired operations to evaluate its performances. Results shown in Table 7.4-1 indicate the FMC system is equally effective in controlling SO_2 emissions from an oil-fired boiler as it was from a coal-fired boiler. As with the coal-fired tests at the Firestone facility, SO_2 emission measurements were taken with a pulsed fluorescent analyzer.[22]

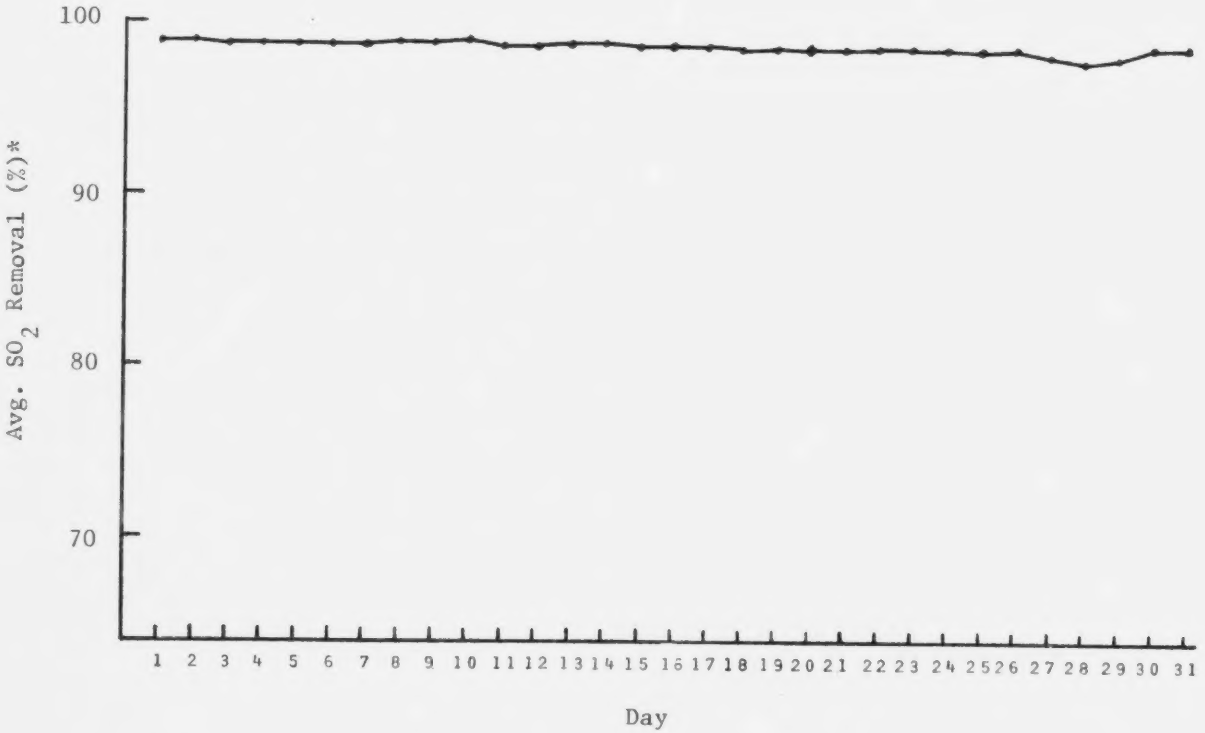
The two examples of 30-day continuous monitoring data (supplied by Dr. Ando) for industrial boiler FGD systems are shown in Figures 7.4-1 and 7.4-2. Figure 7.4-1 is a representation of the data (during October 1978) from Kureha Chemical's sodium throwaway process applied to a 71 MW_e boiler burning 3 percent sulfur heavy oil at the Nishiki plant. This process uses a packed tower to remove SO_2 from the flue gas of an existing boiler (a retrofit FGD installation). One can see from Figure 7.4-1 that the process operated at 90+ percent SO_2 removal for all 31 days. It should be pointed out here that the daily average SO_2 removal is comprised of three readings, one taken at 6:00 a.m., 1:00 p.m., and 9:00 p.m each day. In a recent one-year time frame, from April 1, 1977 to March 31, 1978, both the boiler and the FGD system operated for 8516 hours.[23]

Figure 7.4-2 is a representation of the data (during May 1977) from the Wellman-Lord process applied to a new 135 MW_e installation burning 3 percent sulfur heavy oil at Shindaikyowa Petrochemical's Yokkaichi plant. One can see from Figure 7.4-2 that the process operated at 90+ percent SO_2 removal for all 31 days. While only one reading per day, using an infrared SO_2 analyzer, was taken for the majority of this time, readings were taken every other hour for three consecutive days to check on fluctuations. These bihourly readings showed that there was a maximum deviation in SO_2 removal during any one day of

TABLE 7.4-1. OIL-FIRED EMISSION SOURCE DATA
FIRESTONE TIRE AND RUBBER - FMC DOUBLE-ALKALI FGD SYSTEM[21]

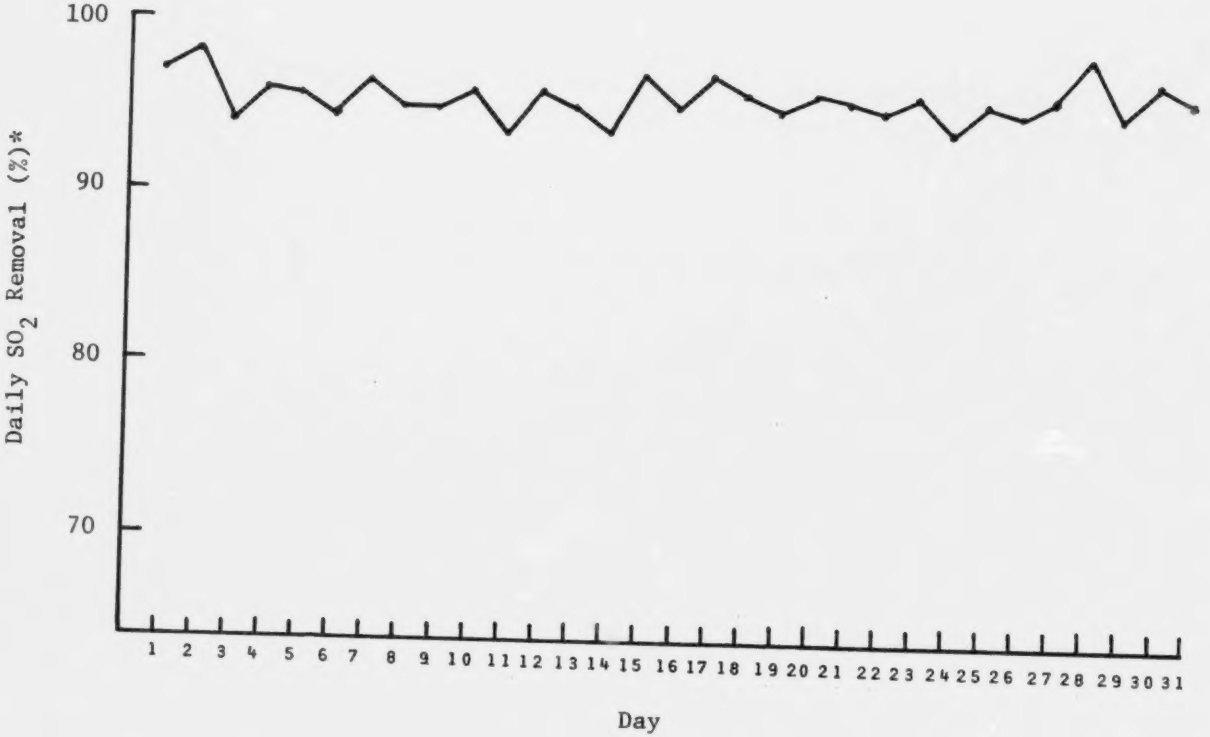
Test No.	Oil feed rate (gas/hr)	Oil heat value kg/kg	SO_2 Inlet		SO_2 outlet		SO_2 control efficiency
			ng/J	lb/10^6 Btu	ng/J	lb/10^6 Btu	
202-1	900	*	938	2.18	32.1	0.07	96.6
202-2	900	*	1075	2.50	29.2	0.07	97.3
202-3	880	*	1085	2.52	26.7	0.06	97.5
202-4	805	*	874	2.03	19.2	0.04	97.8
Average	871	40,741*	993	2.31	26.8	0.06	97.3

*The heat content of the oil burned is nearly constant at this value; individual values were not available.



*Note: The average daily removal is comprised of 3 readings, taken at 8-hour intervals.

Figure 7.4-1. Performance of Kureha Chemical's sodium throwaway process applied to a 71 MW_e boiler burning 3 percent sulfur heavy oil at the Nishiki plant during October 1978.[23]



*Note: The daily removal was taken each day at 1:00 pm. On three consecutive days (May 7,8,9) readings were taken every other hour to gauge amplitude of fluctuations.

Figure 7.4-2. Performance of Wellman-Lord process applied to a 135 MW_e boiler burning 3 percent sulfur heavy oil at Shindaikyowa Petrochemical's Yokkaichi plant during May 1977.[23]

only about three percent. Therefore, the once a day readings for the 31 days of testing should adequately represent the operation of the system.

These good examples of continuous monitoring data are for heavy oil-fired boilers and also for larger boiler sizes than those for the standard industrial boilers of this study; however, the performance of these FGD systems certainly demonstrates their ability to achieve the stringent level of SO_2 control.

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SECTION 8

ADDITIONAL FGD TOPICS

This section provides supplemental evaluations of the candidate control systems previously considered for application to small industrial boilers. Included in this section are evaluations of partial scrubbing for limestone, double alkali, and sodium processes; and evaluations of high sulfur coal limestone applications. The technologies are assessed in light of their economic, energy, and environmental considerations in the same manner as the candidate systems evaluated in Sections 4, 5, and 6 of this report.

Three additional scrubbing cases which treat only part of the flue gas were considered for these processes in order to develop cost and energy correlations for partial scrubbing. The results of material and energy balances are presented in this section for cases treating 75, 50, and 25 percent of the flue gas from coal-fired boilers (3.5% S coal, 90% removal). To study the effect of coal variability on scrubber operations, the costs and energy requirements for a limestone scrubbing system applied to a 5 percent sulful coal are included.

8.1. PARTIAL SCRUBBING

8.1.1. Control Costs

The cost information for the partial scrubbing cases was developed in the same manner as in Section 4 of this report. Material and energy balances were calculated for each additional case and the equipment items discussed in Section 4 were sized according to the material balance results. Equipment cost data obtained from vendors was then used to estimate costs for each of the FGD processes. The base cases for the partial scrubbing comparisons are the 90 percent SO_2 removal cases treating 100 percent of the flue gas from boilers burning 3.5 percent sulfur coal. The calculations in this section were extended to evaluate the impacts of treating 75, 50, and 25 percent of the flue gas from these boilers.

Tables 8.1-1 through 8.1-3 summarize the capital and annualized costs for the partial scrubbing cases. These data are also plotted in Figures 8.1-1 and 8.1-2 for the sodium throwaway process to illustrate the cost sensitivity of partial scrubbing. Throughout this discussion, the sodium throwaway process will be used as the example to illustrate the impacts of partial scrubbing. Tabular results of the impacts of partial scrubbing on the limestone and double alkali processes are provided so that one can verify that the trends shown by the sodium throwaway process are also true for the other FGD processes being evaluated.

For the 58.6 MW_t (200 x 10^6 Btu/hr) applications, the sodium throwaway capital costs ranged from $1,214,000 (100 percent treated) to $580,000 (25 percent treated). Annualized costs ranged accordingly from $1,207,000 to $520,000. Capital costs were reduced roughly by one-half in going from 100 percent scrubbing to 25 percent scrubbing, a reduction by 4 in the amount of SO_2 removal. The same factor is also evident for the annualized costs and simply illustrates the economy of scale for the process. The other processes also show similar economies of scale.

Data are provided to compare the impact of removing 75 percent of the SO_2 from all the gas versus removing 90 percent of the SO_2 from 75 percent of the gas (67.5 percent removal). Results presented in Table 8.1-1 indicate that removing 90 percent of the SO from a portion of the gas (partial scrubbing) is more economical than low efficiency (75%) full scrubbing. This is because a significant portion of the process capital costs (scrubber, fans)

TABLE 8.1-1. SODIUM THROWAWAY PARTIAL SCRUBBING COST SUMMARY

(Eastern 3.5% S Coal)

Boiler Size & Type	Percent of Gas Scrubbed	Percent SO_2 Removal	Total Capital Investment (10^3 \$)	Annualized Costs (10^3 \$)	Annualized Costs Percent increase over uncontrolled
22 MW_t	100	90	730	646	33
(75×10^6 Btu/hr)	100	75	705	596	31
Chaingrate Stoker	75	67.5	590	540	28
	50	45.0	500	460	24
44 MW_t	100	90	1077	1010	33
(150×10^6 Btu/hr)	100	75	1033	917	30
Spreader Stoker	75	67.5	890	822	27
	50	45	720	640	21
	25	22.5	510	440	14
58.6 MW_t	100	90	1214	1207	27
(200×10^6 Btu/hr)	100	75	1160	1080	25
Pulverized Coal	75	67.5	1080	1010	23
	50	45	840	740	17
	75	22.5	580	520	12
118 MW_t	100	90	1911	2063	27
(400×10^6 Btu/hr)	75	67.5	1600	1625	21
Pulverized Coal	50	45	1214	1207	16
	25	22.5	840	740	10

TABLE 8.1-2. DOUBLE ALKALI PARTIAL SCRUBBING COST SUMMARY
(Eastern 3.5% S coal)

Boiler Size and Type	Percent of Gas Scrubbed	Percent SO_2 Removal	Total Capital Investment (10^3\$)	Annualized Costs	
				(10^3\$)	Percent Increase Over Uncontrolled
22 MW_t	100	90	960	625	32
(75×10^6 Btu/hr)	75	67.5	840	530	27
Chaingrate Stoker	50	45	750	460	24
58.6 MW_t	100	90	1422	1053	25
(200×10^6 Btu/hr)	75	67.5	1220	870	20
Pulverized Coal	50	45	1050	695	16
	25	22.5	812	505	12
118 MW_t	100	90	2064	1778	23
(400×10^6 Btu/hr)	75	67.5	1740	1420	18
Pulverized Coal	50	45	1420	1053	14
	25	22.5	1050	695	9

TABLE 8.1-3. LIMESTONE PARTIAL SCRUBBING COST SUMMARY

(Eastern 3.5% S Coal)

Boiler Size & Type	Percent of Gas Scrubbed	Percent SO_2 Removal	Total Capital Investment (10^3 \$)	Annualized Costs	
				(10^3\$)	Percent increase over uncontrolled boilers
22 MW_t	100	90	987	650	34
($75x10^6$ Btu/hr)	100	75	913	593	31
Chaingrate Stoker	75	67.5	860	560	29
	50	45	730	468	24
44 MW_t	100	90	1385	974	32
($150x10^6$ Btu/hr	100	75	1270	865	28
Spreader Stoker	75	67.5	1180	805	26
	50	45	970	630	20
	25	22.5	730	470	15
58.6 MW_t	100	90	1530	1155	26
($200x10^6$ Btu/hr)	100	75	1392	1014	23
Pulverized Coal	75	67.5	1385	960	22
	50	45	1120	735	17
	25	22.5	820	520	12

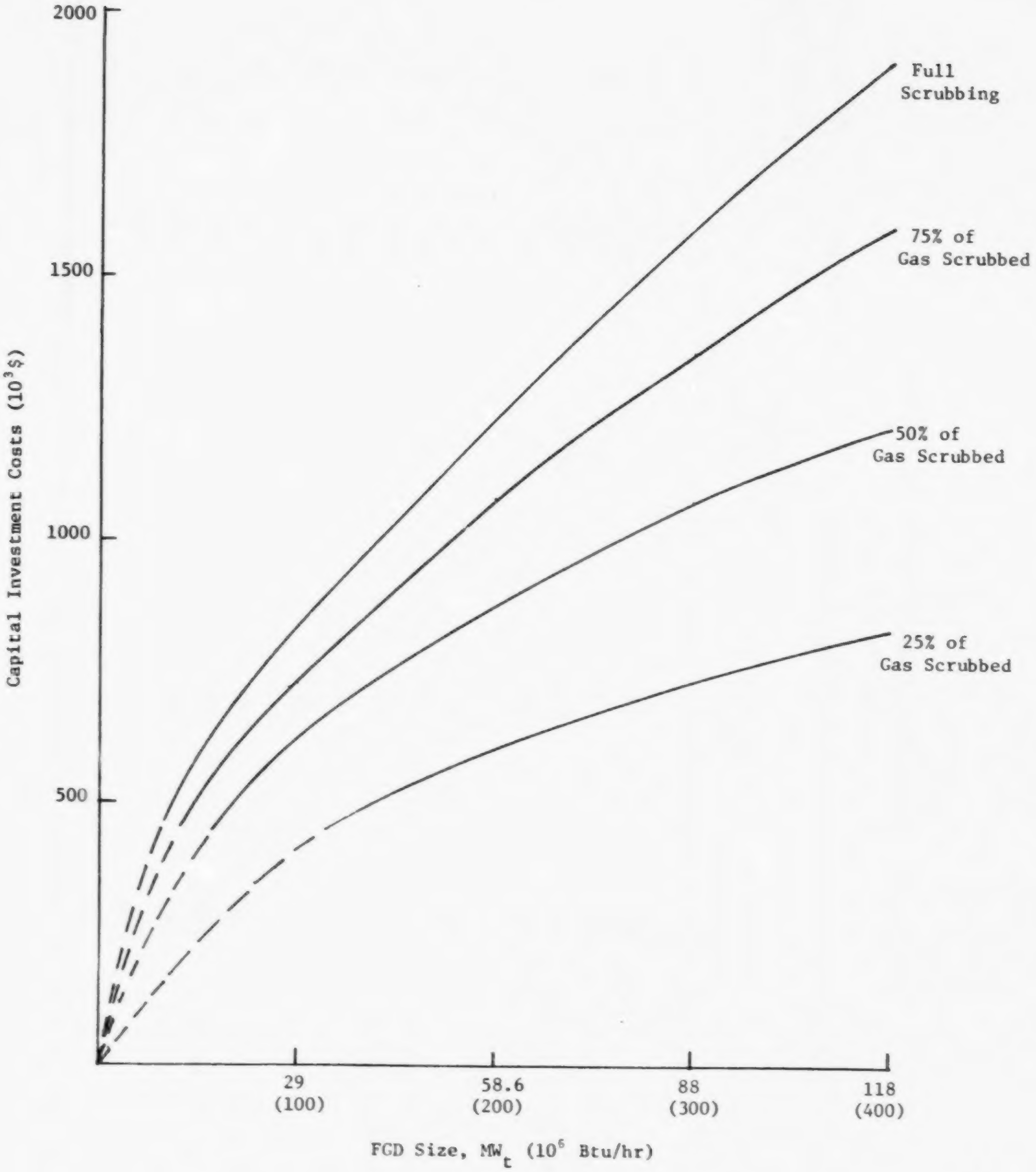


Figure 8.1-1. Sodium Partial Scrubbing Capital Investment Costs (3.5% S Coal, 90% Removal)

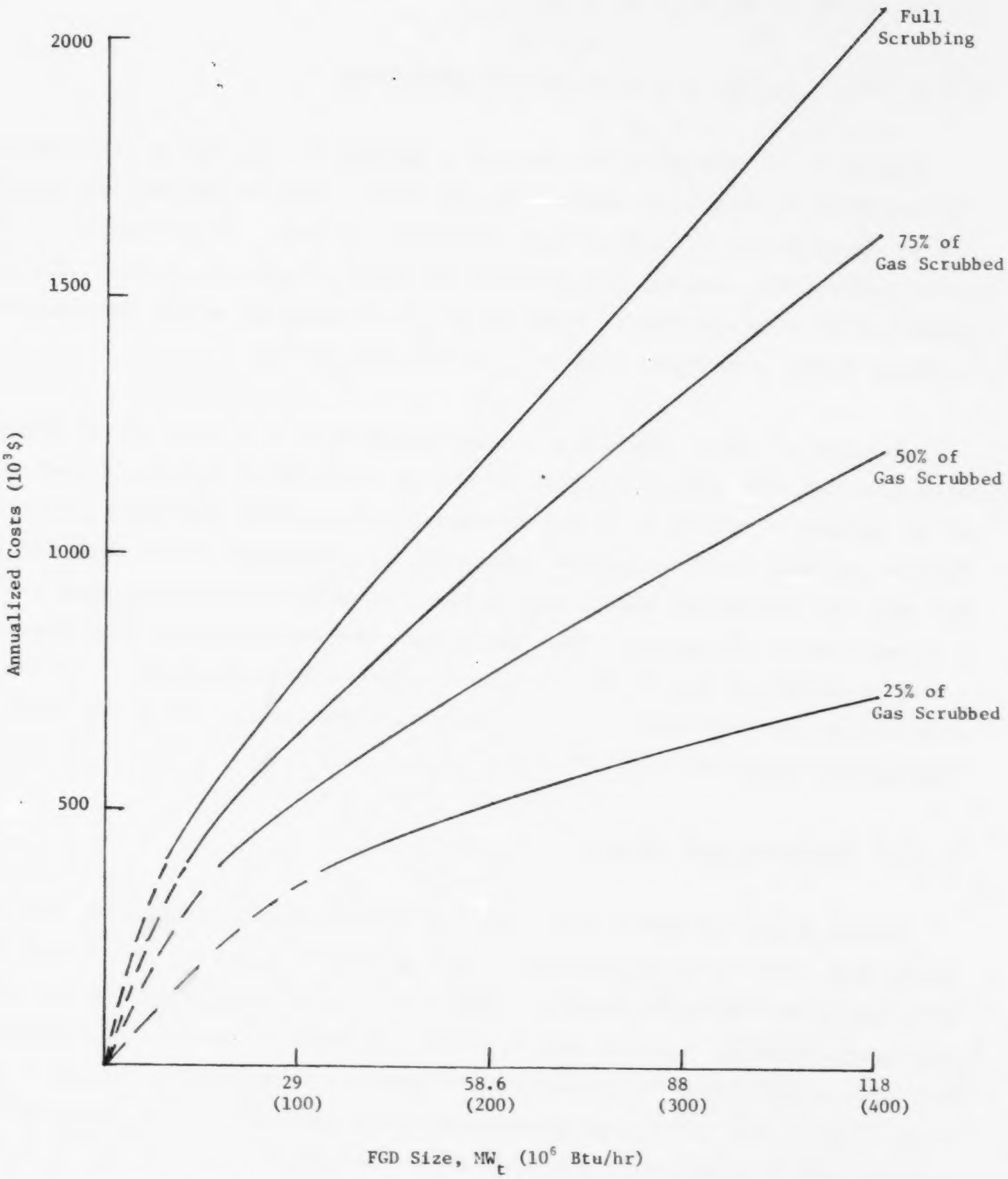


Figure 8.1-2. Sodium Partial Scrubbing Annual Costs (3.5% S Coal, 90% Removal)

are based on gas flow rate and remain constant for a given sized application whereas costs for partial scrubbing processes take advantage of a reduction in the volume of gases to be treated.

8.1.2 Energy Requirements for Partial Scrubbing

Tables 8.1-4 through 8.1-6 present a summary of the energy requirements for the partial scrubbing cases. Figure 8.1-3 presents the partial scrubbing energy requirements for the sodium throwaway process. These cases are based on 90 percent SO_2 removal from boilers burning 3.5 percent S coal. Energy requirements were calculated from the material balances using the same calculation bases presented in Table 5.2-1 of this report.

Results of these calculations show essentially a linear use of energy with size for the various partial scrubbing cases which indicates that there is no economy of scale or energy savings for the sodium throwaway process for the partial scrubbing cases. However, if stack gas reheat were required for the full scrubbing cases, the partial scrubbing cases would show a significant energy advantage. This is because the reheat energy requirements could be saved for the partial scrubbing cases since bypassing 19 percent of the flue gas provides sufficient heat to maintain the stack gas exit temperature above 353°K (175°F).

8.1.3. Environmental Impacts

Tables 8.1-7 through 8.1-9 present a summary of the wastewater and sludge production rates for the partial scrubbing cases. Figure 8.1-4 illustrates the partial scrubbing wastewater production rates graphically for the sodium throwaway process. As with the economic and energy impacts, the environmental impacts are based on 90 percent SO removal from boilers burning 3.5% S coal. Wastewater and sludge production rates were calculated using the material balances found in Appendix A of this report.

TABLE 8.1-4. SODIUM THROWAWAY PARTIAL SCRUBBING ENERGY REQUIREMENTS
(Eastern 3.5% S coal)

Boiler Size and Type	Percent of Gas Scrubbed	Percent SO_2 Removal	Energy Consumption		Percent Increase Over Uncontrolled Boiler
			KW_t	(10^6 Btu/hr)	
22 MW_t	100	90	181	0.6	0.8
(75×10^6 Btu/hr)	100	75	168	0.6	0.8
Chaingrate Stoker	75	67.5	125	0.4	0.5
	50	45	85	0.3	0.4
44 MW_t	100	90	363	1.2	0.8
(150×10^6 Btu/hr)	100	75	337	1.1	0.7
Spreader Stoker	75	67.5	265	0.9	0.6
	50	45	168	0.6	0.4
	25	22.5	85	0.3	0.2
58.6 MW_t	100	90	442	1.5	0.8
(200×10^6 Btu/hr)	100	75	409	1.4	0.7
Pulverized Coal	75	67.5	360	1.2	0.6
	50	45	222	0.8	0.4
	25	22.5	112	0.4	0.2
118 MW_t	100	90	873	3.0	0.8
(400×10^6 Btu/hr)	75	67.5	660	2.2	0.6
Pulverized Coal	50	45	440	1.5	0.4
	25	22.5	220	0.8	0.2

TABLE 8.1-5. DOUBLE ALKALI PARTIAL SCRUBBING ENERGY REQUIREMENTS

(Eastern 3.5% S Coal)

Boiler Size & Type	Percent of Gas Scrubbed	Percent SO_2 Removal	Energy Consumption		Percent increase over uncontrolled boiler
			kW_t	(10^6 Btu/hr)	
22 MW_t	100	90	130	0.4	0.5
(75×10^6 Btu/hr)	75	67.5	90	0.3	0.4
Chaingrate Stoker	50	45	63	0.2	0.3
58.6 MW_t	100	90	307	1.0	0.5
(200×10^6 Btu/hr)	75	67.5	234	0.8	0.4
Pulverized Coal	50	45	157	0.5	0.3
	25	22.5	82	0.3	0.2
118 MW_t	100	90	606	2.1	0.5
(400×10^6 Btu/hr)	75	67.5	460	1.6	0.4
Pulverized Coal	50	45	307	1.0	0.3
	25	22.5	157	0.5	0.1

TABLE 8.1-6. LIMESTONE PARTIAL SCRUBBING ENERGY REQUIREMENTS

(Eastern 3.5% S coal)

Boiler Size & type	Percent of Gas Scrubbed	Percent SO_2 Removal	Energy Consumption		Percent Increase over uncontrolled boiler
			KW_t	(10^6 Btu/hr)	
22 MW_t	100	90	387	1.3	1.7
(75 x 10^6 Btu/hr)	100	75	253	0.9	1.2
Chaingate Staker	75	67.5	290	1.0	1.3
	50	45	195	0.7	0.9
44 MW_t	100	90	777	2.7	1.8
($150x10^6$ Btu/hr)	100	75	507	1.7	1.1
Spreader Stoker	75	67.5	570	1.9	1.3
	50	45	380	1.3	0.9
	25	22.5	195	0.7	0.5
58.6 MW_5	100	90	912	3.1	1.6
($200x10^6$ Btu/hr)	100	75	599	2.1	1.1
Pulverized Coal	75	67.5	770	2.6	1.3
	50	45	510	1.7	0.9
	25	22.5	245	0.8	0.4

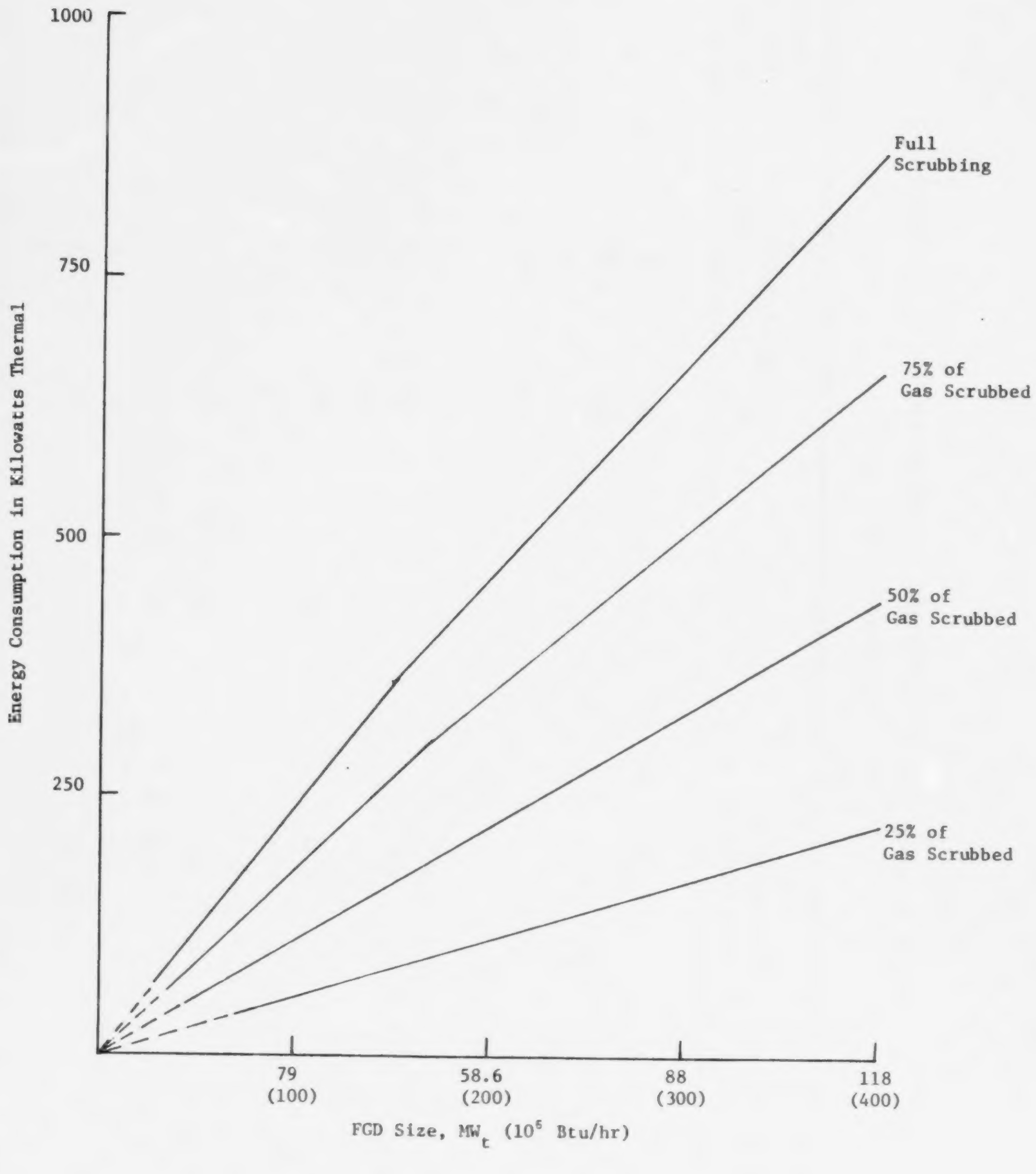


Figure 8.1-3. Sodium Partial Scrubbing Energy Consumption
(3.5% S Coal, 90% removal)

TABLE 8.1-7. SODIUM THROWAWAY PARTIAL SCRUBBING WASTEWATER PRODUCTION RATES (Eastern 3.5% S coal)

Boiler Size & Type	Percent of gas Scrubbed	Percent of SO_2 Removal	Wastewater Production Rate	
			ℓ/sec	(gpm)
22 MW_t	100	90	2.2	34.6
(75 x 10^6 Btu/hr)	100	75	1.8	28.1
Chaingate Stoker	75	67.5	1.5	23.8
	50	45	1.0	16.5
44 MW_t	100	90	4.4	69.4
(150 x 10^6 Btu/hr)	100	75	3.5	55.9
Spreader Stoker	75	67.5	3.2	50.5
	50	45	2.2	34.6
	25	22.5	1.0	16.5
58.6 MW_t	100	90	5.8	92.0
(200 x 10^6 Btu/hr)	100	75	5.3	83.4
Pulverized Coal	75	67.5	4.4	69.4
	50	45	2.9	45.3
	25	22.5	1.4	22.2
118 MW_t	100	90	11.7	185.0
(400 x 10^6 Btu/hr)	75	67.5	8.8	139.0
Pulverized Coal	50	45	5.8	92.0
	25	22.5	2.9	45.3

TABLE 8.1-8. DOUBLE ALKALI PARTIAL SCRUBBING SLUDGE PRODUCTION RATES (Eastern 3.5% S coal)

Boiler Size and Type	Percent of Gas Scrubbed	Percent SO_2 Removed	Sludge Production Rates	
			g/sec	(lb/hr)
22 MW_t	100	90	200	1584
(75 x 10^6Btu/hr)	75	67.5	145	1150
Chaingrate Stoker	50	45	95	750
58.6 MW_t	100	90	506	4012
(200 x 10^6Btu/hr)	75	67.5	391	3100
Pulverized Coal	50	45	259	2050
	25	77.5	126	1000
118 MW_t	100	90	1066	8450
(400 x 10^6Btu/hr)	75	67.5	788	6250
Pulverized Coal	50	45	506	4012
	75	77.5	259	2050

TABLE 8.1-9. LIMESTONE PARTIAL SCRUBBING SLUDGE PRODUCTION RATES

(Eastern 3.5% S Coal)

Boiler Size & Type	Percent of Gas Scrubbed	Percent SO_2 Removal	Sludge Production Rates	
			g/sec	(lb/hr)
22 MW_t	100	90	227	1796
(75×10^6 Btu/hr)	100	75	190	1508
Chaingrate Stoker	75	67.5	169	1340
	50	45	108	860
44 MW_t	100	90	457	3624
(150×10^6 Btu/hr)	100	75	381	3020
Spreader Stoker	75	67.5	343	2720
	50	45	220	1750
	25	22.5	108	860
58.6 MW_t	100	90	607	4812
(200×10^6 Btu/hr)	100	75	576	4568
Pulverized Coal	75	67.5	458	3630
	50	45	300	2380
	25	22.5	148	1170

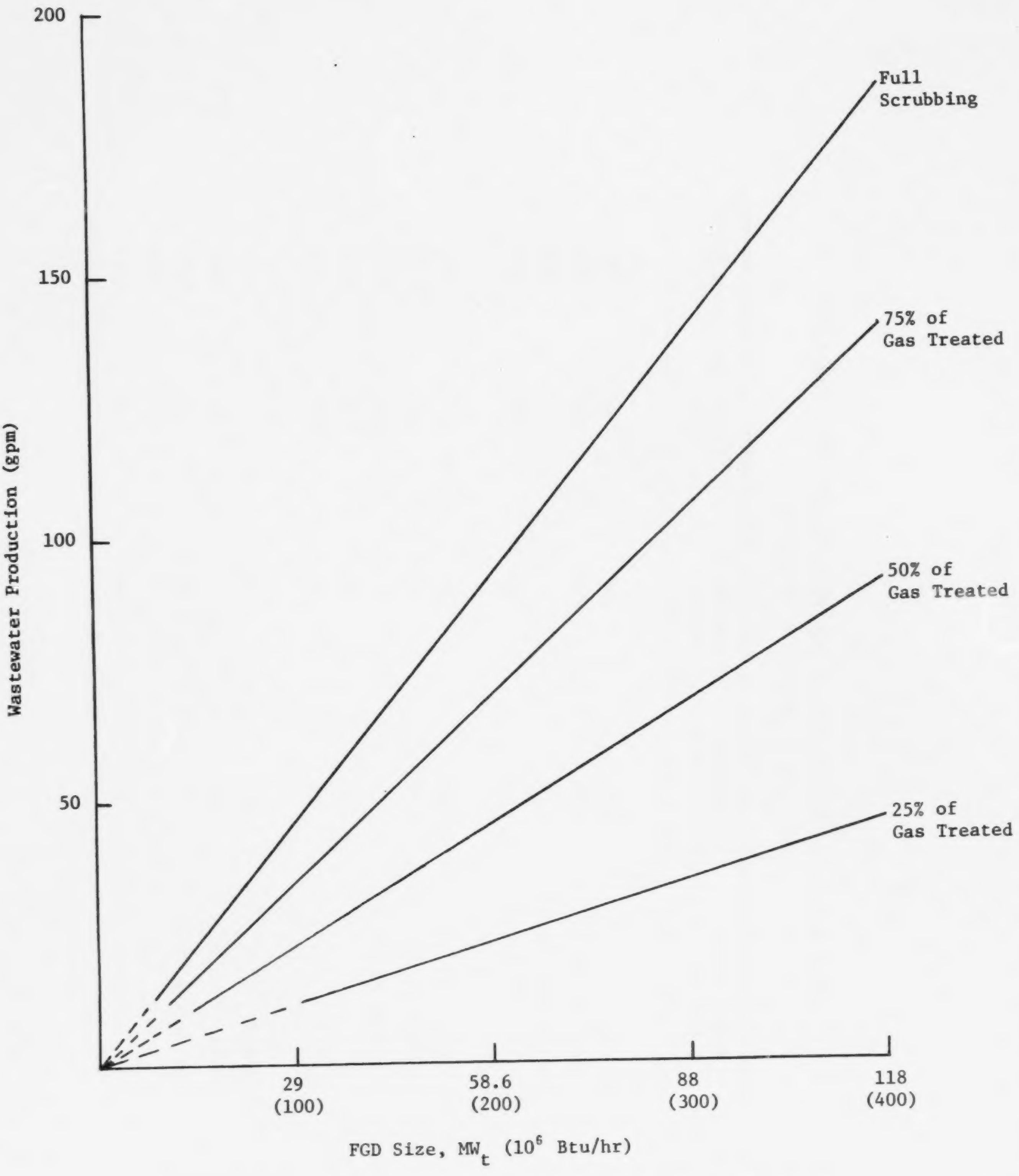


Figure 8.1-4. Sodium Partial Scrubbing Wastewater Production (3.5% S Coal, 90% removal)

Results of these calculations show a linear production of waste materials with size which indicates that there is no economy of scale or environmental benefit resulting from partial scrubbing. This is because the waste production rates vary linearly with the amount of SO_2 removal; that is, a reduction by 50 percent in the SO_2 being removed will result in a 50 percent reduction in the waste production rate for all the throwaway processes.

8.2 LIMESTONE SCRUBBING OF FLUE GAS FROM FIVE PERCENT SULFUR COAL

The purpose of this section is to further examine the effects of coal sulfur content on the capital and annualized costs of a limestone scrubber. Material and energy balances for this case were performed on a limestone scrubber treating the flue gas from a 58.6 MW (200×10^6 Btu/hr) boiler burning 5.0% sulfur coal. Table 8.2-1 presents the bases for the material and energy balance for this case.

TABLE 8.2-1 ASSUMPTIONS AND BASES FOR MATERIAL AND ENERGY BALANCES

Process parameters	Limestone process
L/G, ℓ/m^3 (gal/10^3 acf)[a]	13.3 (100)
Particulate Removal[b]	99 Percent Upstream of Scrubber
Stoichiometry[a] (moles sorbent/mole sorbed SO_2)	1.2
Gas Pressure Drop Pa (in H_2O)[c]	214 (17)
Pump Discharge Pressure Pa(Psi)[b]	5227 (15)
Pumping Height M (ft)[b]	6 (20)
Stack Gas Reheat °C (°F)[b]	28 (50)

[a] Based on process data presented in Section 2

[b] Radian assumption

[c] Based on TVA empirical relationship (Reference 4)

8.2.1 Control Costs

Table 8.2-2 tabulates the cost data for three cases relating capital and annualized costs to variations in sulfur content of coal. Figure 8.2-1 illustrates this relationship. As shown in the plots there is a slight flattening of the capital cost curve as higher sulfur coals are burned. This is due in part to economy of scale coming into effect for the raw material handling facilities. Capital costs increase about 27 percent in going from a 0.6 to 3.5 percent sulfur coal. From 3.5 to 5.0 percent sulfur coal, the capital costs increase by an additional 10 percent. Increased sulfur content to 5 percent should affect the capital costs of other FGD systems in the same manner as for limestone systems. Therefore, the relative ranking of the processes with regard to capital cost are not expected to change at the 5 percent coal level. The increased slope in the annualized cost curve is due to the increased solids disposal cost for the high sulfur coal systems. The relative ranking of FGD processes shown in Section 4 should also not change with regard to annualized costs for the high sulfur coals. Consequently, the sodium throwaway process would still be the most costly for the high sulfur applications.

8.2.2 Control Energy Requirements

An approximate 19 percent increase in energy use is required to go from 3.5 to 5 percent sulfur coal in a 58.6 MW (200×10^6 Btu/hr) boiler assuming 90 percent SO_2 removal. The energy requirements are calculated using the same bases as given in Section 5 of the report. The energy increase is primarily due to increased feed preparation and liquid pumping rates (increased sorbent). Table 8.2-3 shows the energy requirements for a limestone control system treating the flue gas from a 58.6 MW (200×10^6 Btu/hr) boiler burning coals with three different sulfur levels. The other processes show less of an energy impact with changes in coal sulfur content as discussed in Section 5.

TABLE 8.2-2. COST SUMMARY FOR LIMESTONE SCRUBBING

Boiler Size	Coal Type	SO_2 Removal level	Total Capital Investment (10^3\$)	Annualized Costs (10^3\$)	Annualized Costs Percent Increase Over Uncontrolled Use
58.6 MW_t (200x10^6 Btu/hr)	0.6% S	90	1201	634	14
58.6 MW_t (200x10^6 Btu/hr)	3.5% S	90	1530	1155	26
58.6 MW_t (200x10^6 Btu/hr)	5.0% S	90	1689	1426	32

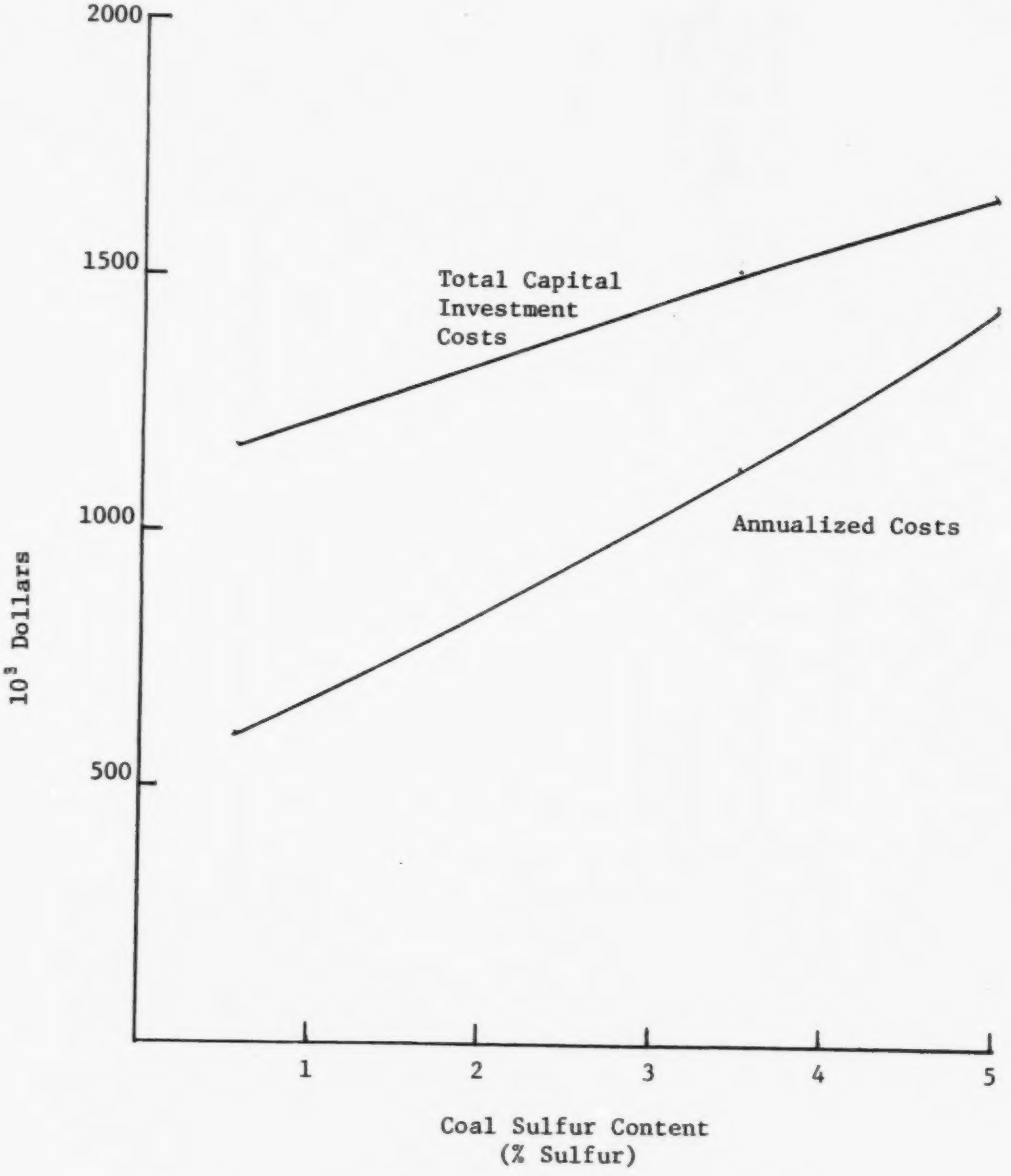


Figure 8.2-1. Capital and Annualized Costs for Limestone Scrubbing *versus* Coal Sulfur Content

TABLE 8.2-3. ENERGY SUMMARY FOR LIMESTONE SYSTEMS

Boiler Size	Coal Type	SO_2 Removal	Energy KW_t	Consumption 10^6 Btu/hr	Percent Increase Over Uncontrolled
58.6 MW_t (200×10^6 Btu/hr)	0.6% S	90	711	2.4	1.2
58.6 MW_t (200×10^6 Btu/hr)	3.5% S	90	912	3.1	1.6
58.6 MW_t (200×10^6 Btu/hr)	5.0% S	90	1076	3.7	1.9

8.2.3. Environmental Impact of Burning Five Percent Sulfur Coal

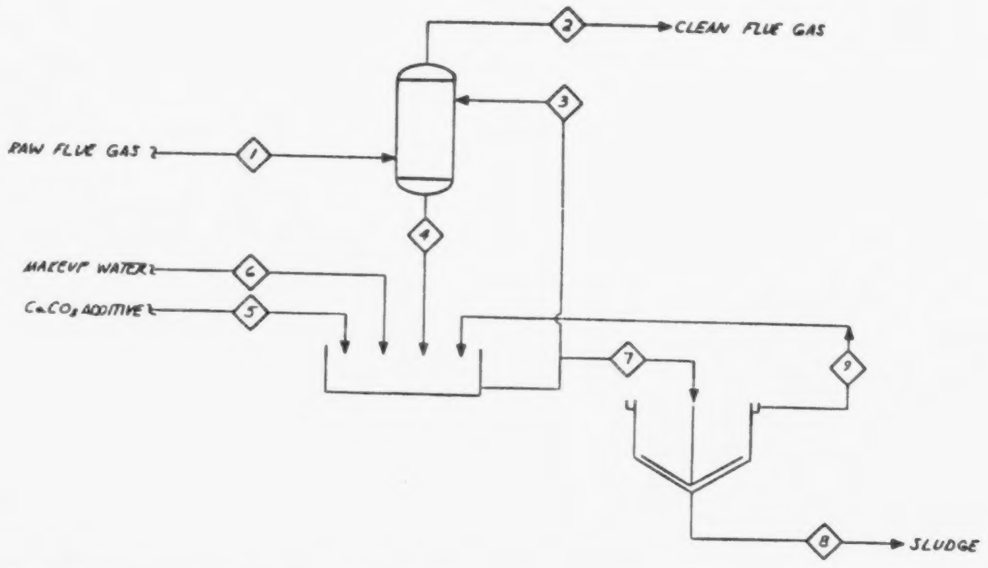
The solid waste production rates for burning high sulfur coal in a limestone process are shown in Table 8.2-4. This table shows a proportional increase in solid waste with coal sulfur content which is expected since waste production rates vary linearly with the amount of SO_2 removed. This same trend is also true with regard to waste production rates for the other FGD processes.

TABLE 8.2-4. SOLID WASTE SUMMARY FOR LIMESTONE SYSTEMS

Boiler Size	Coal Type	SO_2 Removal	Waste Study Production g/sec	Waste Study Production (lb/hr)
58.6 MW_t (200x10^6 Btu/hr)	0.6% S	90	139	1106
58.6 MW_t (200x10^6 Btu/hr)	3.5% S	90	607	4812
58.6 MW_t (200x10^6 Btu/hr)	5.0% S	90	867	6879

APPENDIX A

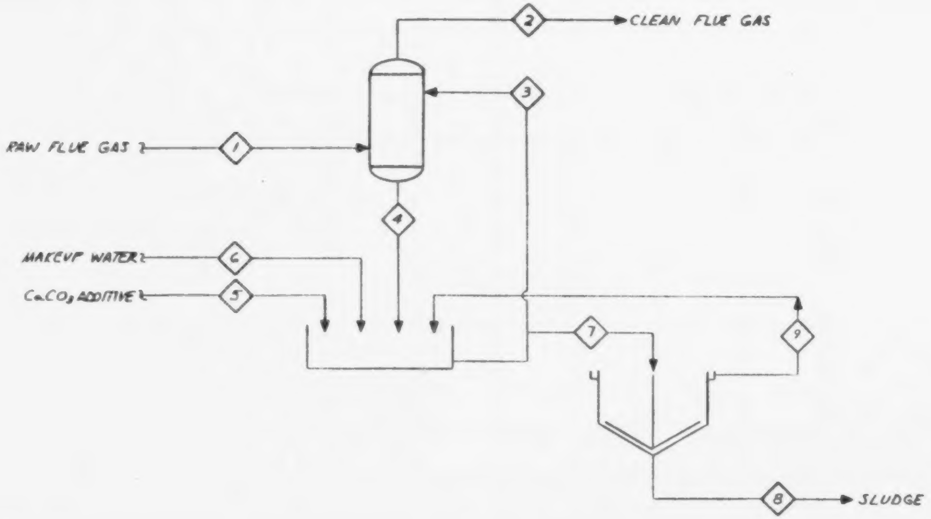
MATERIAL BALANCE CALCULATIONS



LIMESTONE FGD PROCESS

BOILER SIZE: 30×10^6 BTU-Hr.
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%

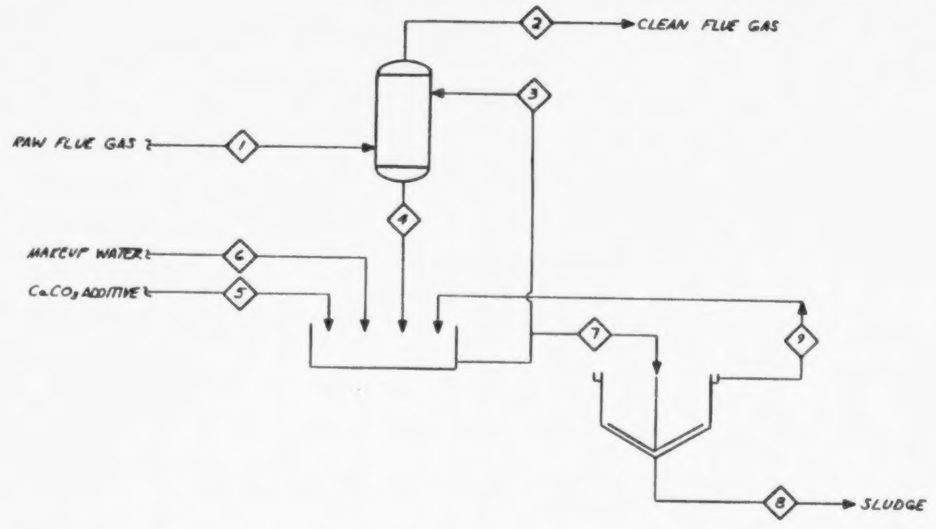
STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb./hr.)	35,810	38,336	575,870	573,360	281	2998	3960	712	2942
(GPM)	—	—	1037.3	1032	—	6.0		0.7	5.9
TEMPERATURE (°F)	400	125							
GAS FLOW RATE									
(lb./hr.)	35,810	38,336							
(lb MOLES/hr.)	1234	1379							
(ACFM)	12,900	9800							
(SCFM)	7800	8711							
LIQUID PHASE FLOW RATE									
(lb./hr.)			518,270	515,760		2998	3564	356	2902
SOLID PHASE FLOW RATE									
(lb./hr.)			57,600	57,600	281		396	356	40



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb./hr.)	35,810	38,336	432,771	430,122	265	2998	3736	672	2776
(GPM)	—	—	779.3	774	—	6.0	6.75	0.7	5.5
TEMPERATURE (°F)	400	125	—	—	—	—	—	—	—
GAS FLOW RATE									
(lb./hr.)	35,810	38,336							
(lb MOLES/hr.)	1234	1379							
(ACFM)	12,900	9800							
(SCFM)	7800	8711							
LIQUID PHASE FLOW RATE (lb./hr.)			389,494	386,845		2998	3363	336	2738
SOLID PHASE FLOW RATE (lb./hr.)			43,277	43,277	265		374	336	38

LIMESTONE FGD PROCESS

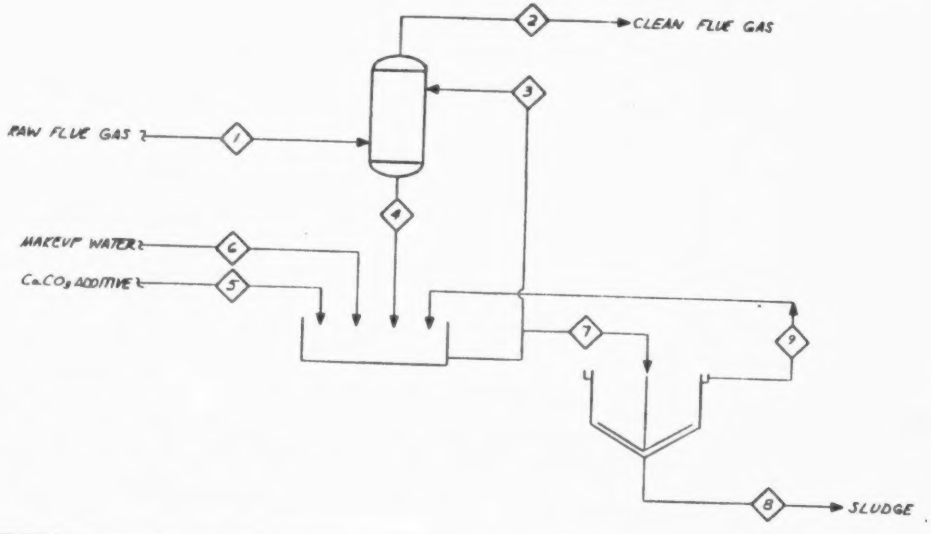
BOILER SIZE: 30×10^6 BTU/hr.
COAL TYPE: EASTERN
SO_2 REMOVAL: 85%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb./hr.)	35,810	38,336	289,494	286,845	234	2949	3300	594	2452
(GPM)	—	—	521.3	516	—	5.9	5.9	0.6	4.8
TEMPERATURE (°F)	400	125	—	—					
GAS FLOW RATE									
(lb./hr.)	35,810	38,336							
(lb. MOLES/hr.)	1234	1379							
(ACFM)	12900	9800							
(SCFM)	7800	8711							
LIQUID PHASE FLOW RATE (lb./hr.)			260545	257,896		2949	2970	297	2419
SOLID PHASE FLOW RATE (lb./hr.)			28949	28949	234		330	297	33

LIMESTONE FGD PROCESS

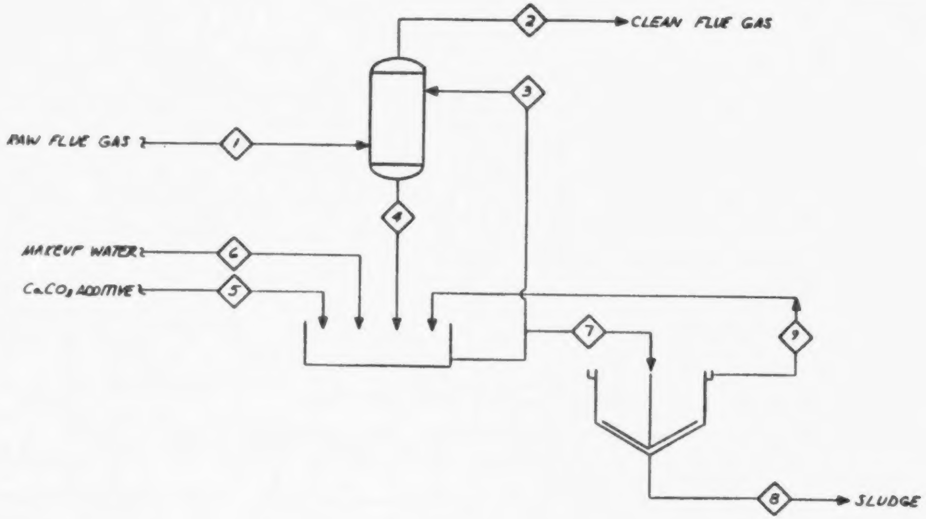
BOILER SIZE: 30×10^6 BTU/hr.
COAL TYPE: EASTERN
SO_2 REMOVAL: 75 %



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 FEED	8 SLUDGE	9 RETURN
TOTAL FLOW RATE									
(lb./hr.)	89663	96076	1.44×10^6	1.43×10^6	708		9980		7413
(GPM)	—	—	2597	2584	—		18		14.6
TEMPERATURE (°F)	400	125							
GAS FLOW RATE									
(lb./hr.)	89663	96,076							
(lb MOLES/hr.)	3091	3456							
(ACFM)	32,300	24559							
(SCFM)	19,530	21830							
LIQUID PHASE FLOW RATE									
(lb./hr.)			1.3×10^6	1.29×10^6			8982		7313
SOLID PHASE FLOW RATE									
(lb./hr.)			144,000	144,000	708		998	898	100

LIMESTONE FGD PROCESS

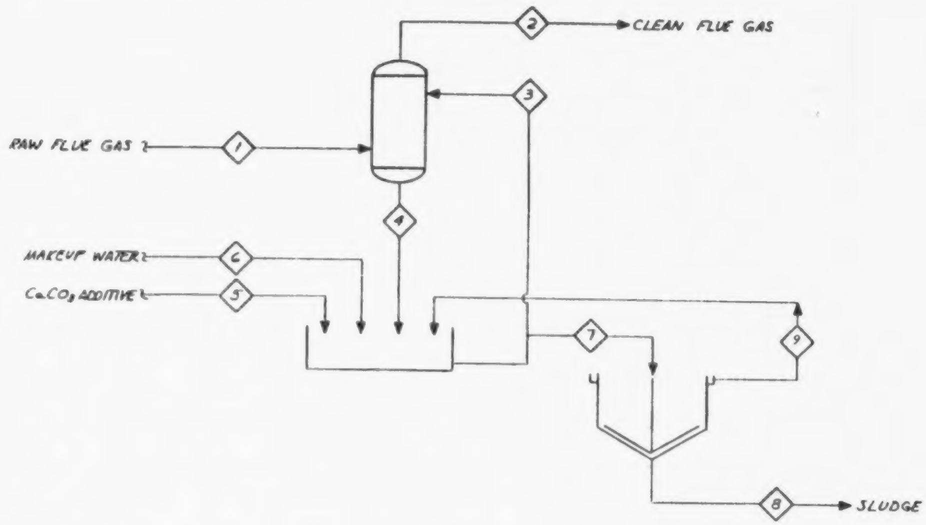
BOILER SIZE: 75×10^6 BTU/hr.
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb./hr.)	89,603	96076	724,931	718235	594	7447	8375	1508	6221
(GPM)	—	—	1305	1292		14.9	15	1.5	12.2
TEMPERATURE (°F)	400	125	—						
GAS FLOW RATE									
(lb./hr.)	89,603	96076							
(lb MOLES/hr.)	3091	3456							
(ACFM)	32,300	24559							
(SCFM)	19530	21830							
LIQUID PHASE FLOW RATE (lb./hr.)			652439	645,742		7447	7537	754	6137
SOLID PHASE FLOW RATE (lb./hr.)			72493	72493	594		837	754	83

LIMESTONE FGD PROCESS

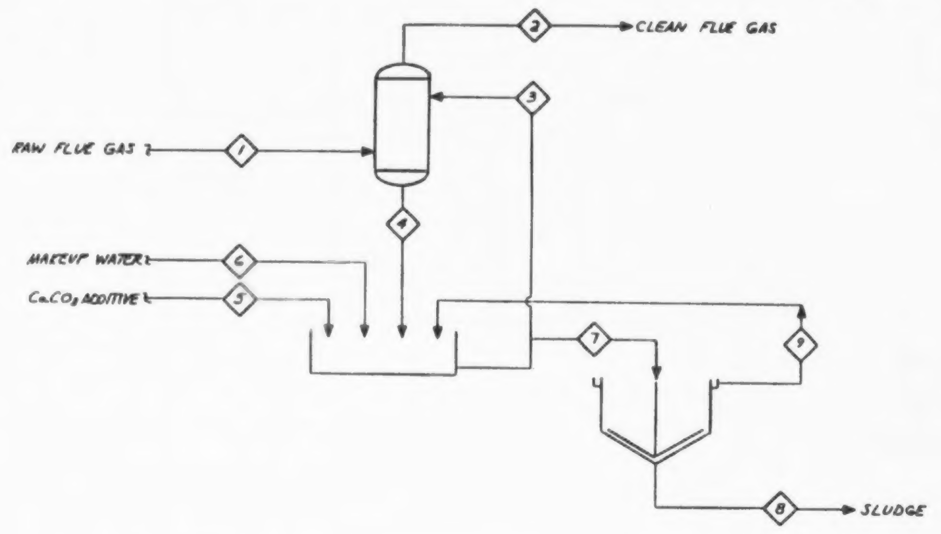
BOILER SIZE: 75 x 10^6 BTU/hr.
COAL TYPE: EASTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb/hr)	179,882	192,765	2.89×10^6	2.89×10^6	1428	15193	20130	3624	14953
(GPM)	—	—	5211	5184		30.4	36.2	3.6	29.5
TEMPERATURE (°F)	400	125							
GAS FLOW RATE									
(lb/hr)	179,882	192765							
(lb MOLES/hr)	6202	6934							
(ACFM)	64800	49,774							
(SCFM)	39180	43,799							
LIQUID PHASE FLOW RATE (lb/hr)			2.6×10^6	2.59×10^6		15193	18117	1812	14752
SOLID PHASE FLOW RATE (lb/hr)			289,373	289,373	1428		2013	1812	201

LIMESTONE FGD PROCESS

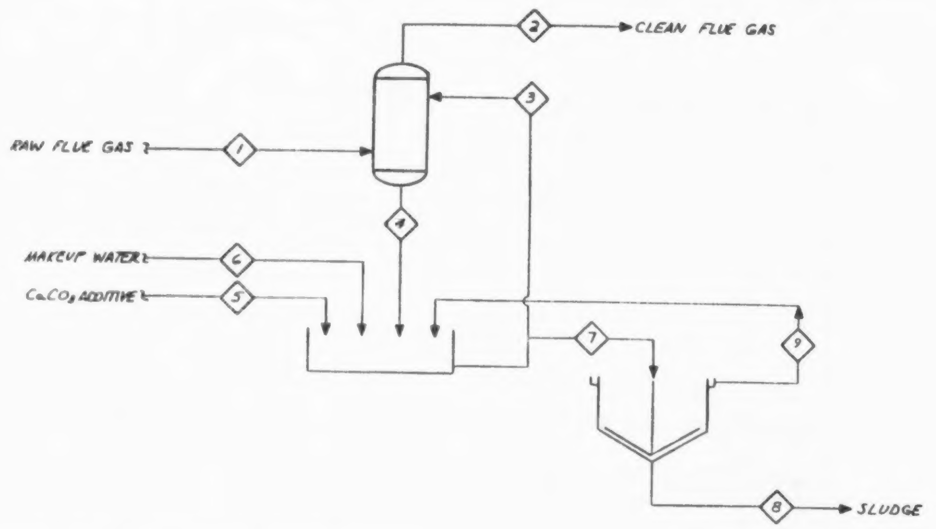
BOILER SIZE: 150×10^6 BTU/hr.
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb/hr)	179,882	192,821	1.45×10^6	1.44×10^6	1188	14894	16750	3018	13161
(GPM)	—	—	2619	2592	—	29.8	30.2	3.0	25.5
TEMPERATURE (°F)	400	125							
GAS FLOW RATE									
(lb/hr)	179,882	192821							
(lb MOLES/hr)	6202	6936							
(ACFM)	64800	49288							
(SCFM)	39180	43812							
LIQUID PHASE FLOW RATE									
(lb/hr)			1.31×10^6	1.29×10^6		15744	15075	1256	12742
SOLID PHASE FLOW RATE									
(lb./hr)			145,430	145,430	1188		1675	1256	419

LIMESTONE FGD PROCESS

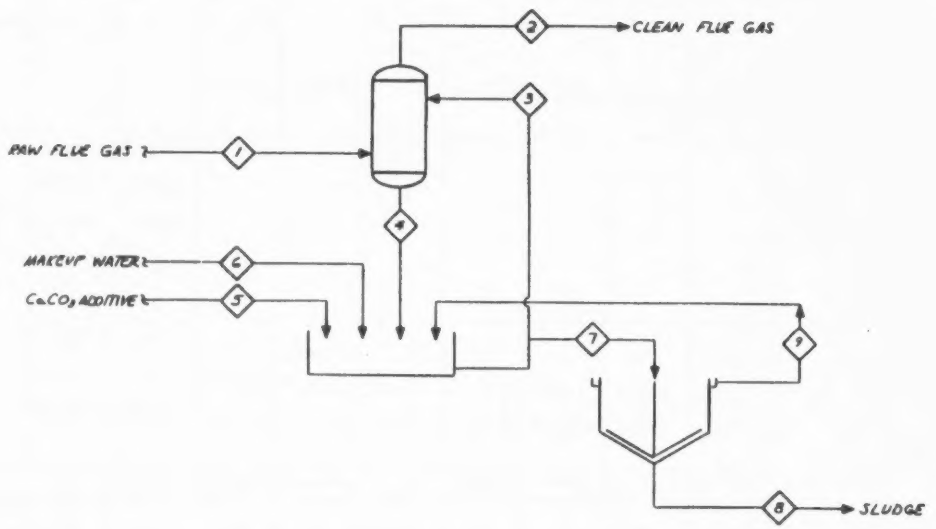
BOILER SIZE: 150 × 10^6 BTU/hr.
COAL TYPE: EASTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb/hr)	207640	222489	3.33×10^6	3.32×10^6	1896	17892	26730	4812	17450
(GPM)	—	—	6015	5984	—	35.8	48.1	4.8	34.3
TEMPERATURE (°F)	400	125	—	—	—	—	—	—	—
GAS FLOW RATE									
(lb/hr)	207640	222489							
(lb MOLES/hr)	7160	8003							
(ACFM)	74800	66871							
(SCFM)	45228	50552							
LIQUID PHASE FLOW RATE									
(lb/hr)			3.00×10^6	2.99×10^6		17892	24057	2406	17183
SOLID PHASE FLOW RATE									
(lb/hr)			334000	334000	1896		2673	2406	267

LIMESTONE FGD PROCESS

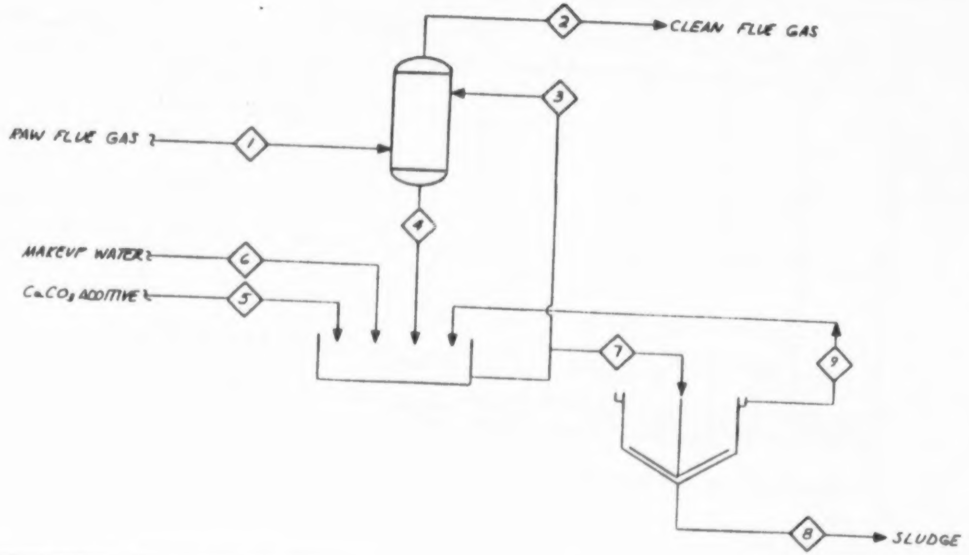
BOILER SIZE: 200×10^6 Btu/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 Clarifier FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb/hr)	207640	222489	2.51×10^6	2.49×10^6	1800	17,792	25,380	9568	18852
(GPM)	—	—	4519	4988	–	35.6	45.7	4.6	37.2
TEMPERATURE (°F)	400	125	–	–	–	–	–	–	–
GAS FLOW RATE									
(lb/hr)	207640	222489							
(lb MOLES/hr)	7160	8003							
(ACFM)	74,800	56,871							
(SCFM)	45,228	50,552							
LIQUID PHASE FLOW RATE									
(lb/hr)			2.26×10^6	2.24×10^6		17,792	22,840	2284	18,592
SOLID PHASE FLOW RATE									
(lb./hr)			250950	250950	1800		2538	2284	254

LIMESTONE FGD PROCESS

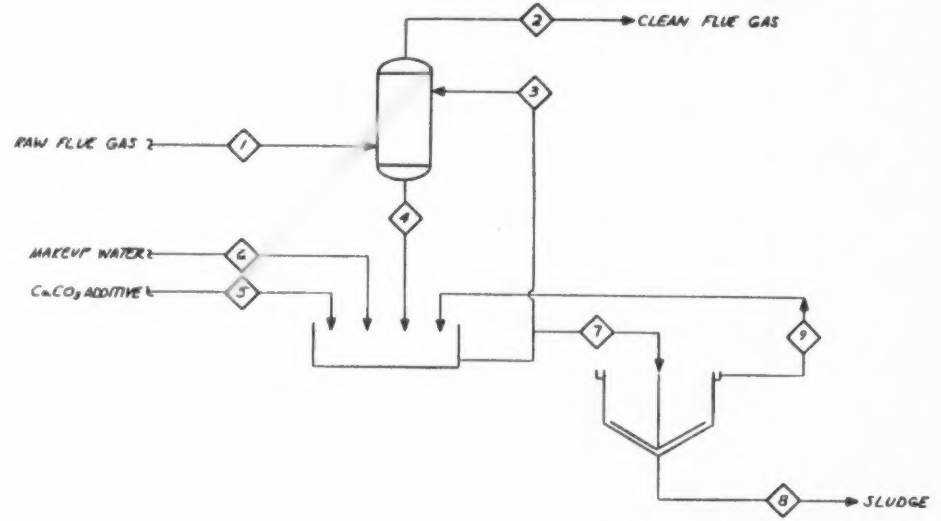
BOILER SIZE: 200×10^6 BTU/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 85%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb./hr.)	207640	222561	1.67×10^6	1.66×10^6	1584	17,493	22330	4020	16587
(GPM)	—	—	3023	2992	—	35.0	40.2	4.0	32.7
TEMPERATURE (°F)	400	125	—	—	—	—	—	—	—
GAS FLOW RATE									
(lb./hr.)	207640	222561							
(lb MOLES/hr.)	7160	8005							
(ACFM)	74800	56885							
(SCFM)	45228	50564							
LIQUID PHASE FLOW RATE									
(lb./hr.)			1.51×10^6	1.49×10^6		17,493	20097	2010	16364
SOLID PHASE FLOW RATE									
(lb./hr.)			167870	167870	1584		2233	2010	223

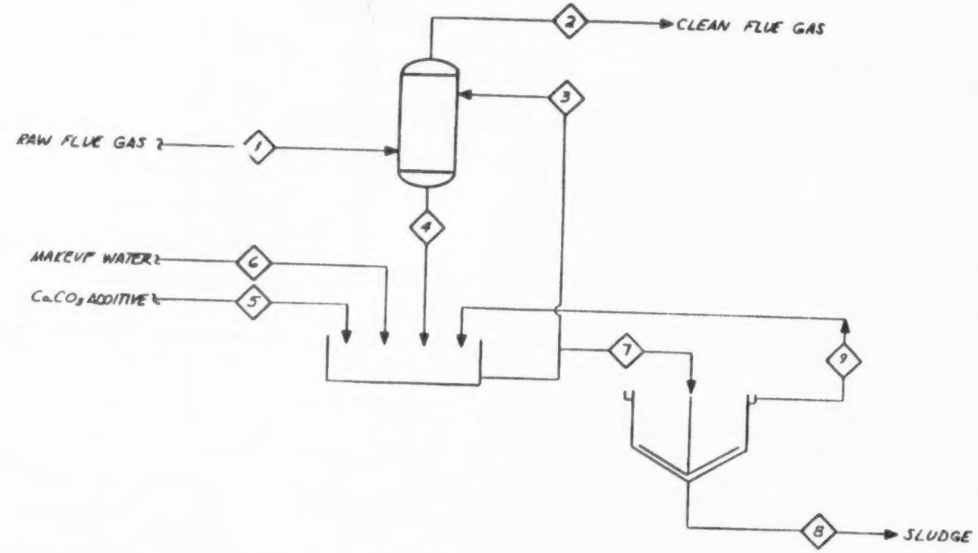
LIMESTONE FGD PROCESS

BOILER SIZE: 200×10^6 BTU/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb/hr.)	36,840	38,836	488,151	486139	60	2399	913	166	676
(GPM)	—	—	879	875	—	4.8	1.6	0.2	1.3
TEMPERATURE (°F)	350	125	—	—	—	—	—	—	—
GAS FLOW RATE									
(lb/hr.)	36840								
(lb MOLES/hr.)	1270	1397							
(ACFM)	12,500	9927							
(SCFM)	8025	8824							
LIQUID PHASE FLOW RATE									
(lb/hr)			439336	437,325		2399	822	83	668
SOLID PHASE FLOW RATE									
(lb./hr.)			48814	48814	60		91	83	8

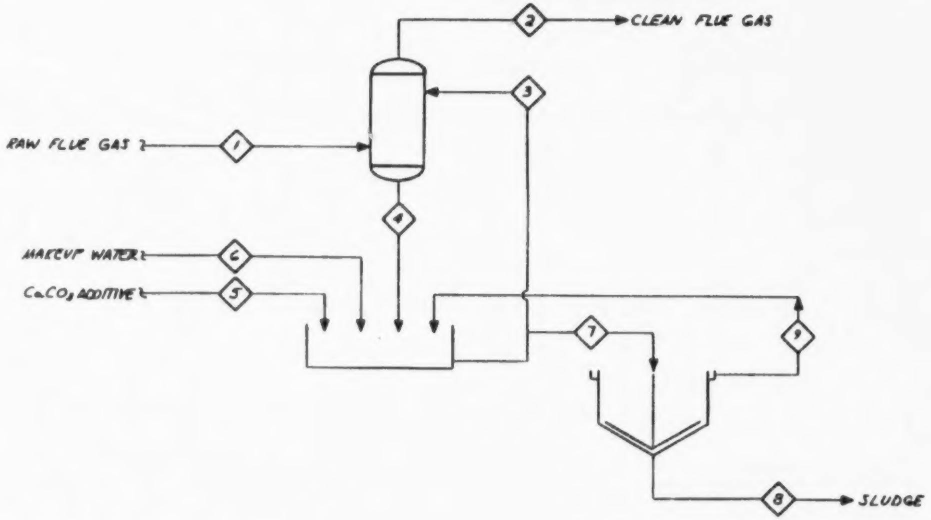
LIMESTONE FGD PROCESS
BOILER SIZE: 30×10^6 BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb./hr.)	36840	38,836	384,133	382,025	56	2384	860	154	640
(GPM)	—	—	691.7	687.5	—	4.7	1.5	0.15	1.25
TEMPERATURE (°F)	350	125	—						
GAS FLOW RATE									
(lb./hr.)	36840	38,836							
(lb MOLES/hr.)	1270	1397							
(ACFM)	12,500	9927							
(SCFM)	8025	8824							
LIQUID PHASE FLOW RATE									
(lb./hr.)			345,720	343612		2384	774	77	631
SOLID PHASE FLOW RATE									
(lb./hr.)			38413	38413	52		86	77	9

LIMESTONE FGD PROCESS

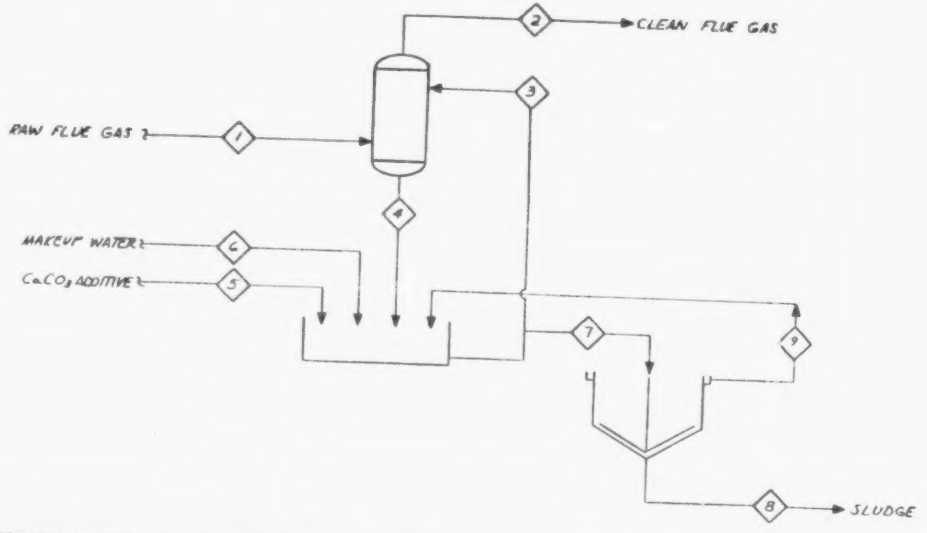
BOILER SIZE: 30 × 10^6 BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 85 %



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb./hr.)	36840	38836	280221	277,922	50	2369	770	138	573
(GPM)	—	—	504.6	500	—	4.7	1.4	0.14	1.15
TEMPERATURE (°F)	350	125							
GAS FLOW RATE									
(lb./hr.)	36840	38,836							
(lb MOLES/hr.)	1270	1397							
(ACFM)	12500	9927							
(SCFM)	8025	8824							
LIQUID PHASE FLOW RATE									
(lb./hr.)			252,199	249,900		2369	693	69	565
SOLID PHASE FLOW RATE									
(lb./hr.)			28,022	28,022	50		77	69	8

LIMESTONE FGD PROCESS

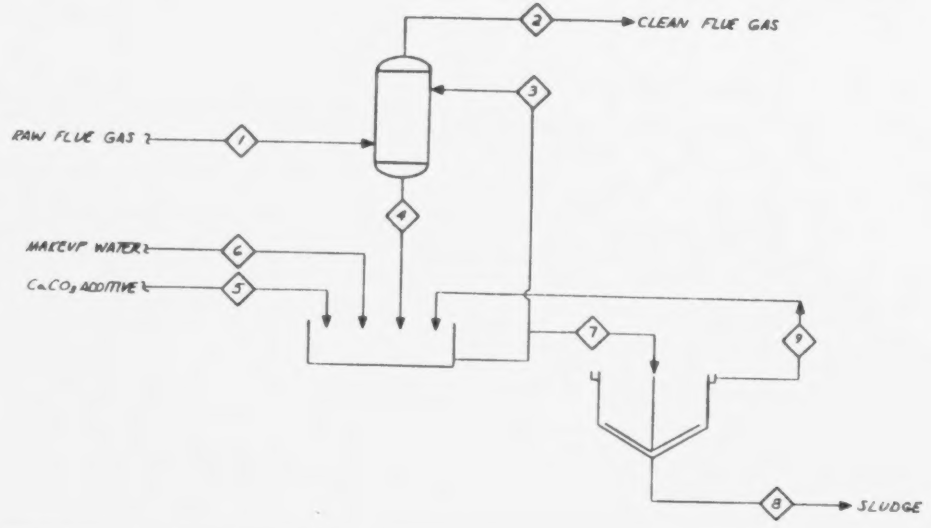
BOILER SIZE: 30×10^6 BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb./hr.)	92840	97856	1.22×10^6	1.21×10^6	150	5947	2300	414	1708
(GPM)	—	—	2215	2205	—	11.9	4.2	0.4	3.4
TEMPERATURE (°F)	350	125							
GAS FLOW RATE									
(lb./hr.)	92840	97856							
(lb MOLES/hr.)	3201	3520							
(ACFM)	31400	25,013							
(SCFM)	20,222	22,234							
LIQUID PHASE FLOW RATE									
(lb./hr.)			1.10×10^6	1.09×10^6		5947	2070	207	1685
SOLID PHASE FLOW RATE									
(lb./hr.)			122,500	122,500	150		230	207	23

LIMESTONE FGD PROCESS

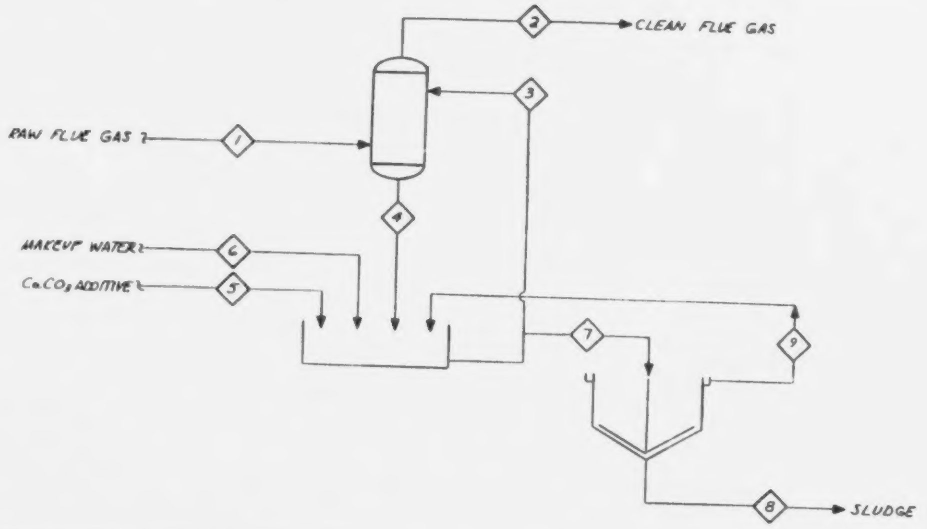
BOILER SIZE: 75×10^6 BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 FEED	8 SLUDGE	9 RETURN
TOTAL FLOW RATE									
(lb./hr.)	92840	97856	703884	698,136	125	5897	1920	346	1427
(GPM)	—	—	1267	1256	—	11.8	3.5	0.3	2.9
TEMPERATURE (°F)	350	125							
GAS FLOW RATE									
(lb./hr.)	92840	97856							
(lb MOLES/hr.)	3201	3520							
(ACFM)	31400	25013							
(SCFM)	20222	22234							
LIQUID PHASE FLOW RATE									
(lb./hr.)			633496	627,748		5897	1728	173	1408
SOLID PHASE FLOW RATE									
(lb./hr.)			70388	70388	125		192	173	19

LIMESTONE FGD PROCESS

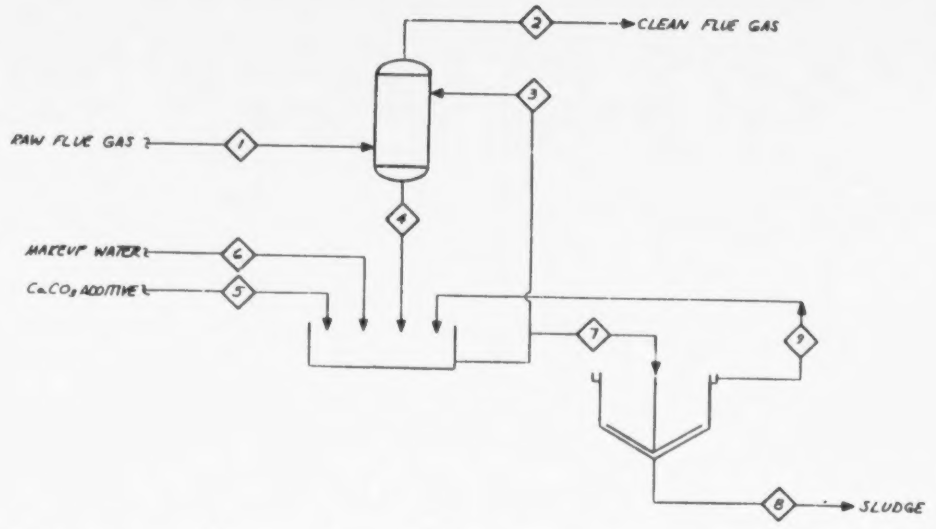
BOILER SIZE: 75×10^6 BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb/hr)	185092	195086	2.45×10^6	2.44×10^6	300	11,895	4610	830	3425
(GPM)	—	—	4416	4396	—	23.8	8.3	0.8	6.8
TEMPERATURE (°F)	350	125							
GAS FLOW RATE									
(lb/hr)	185092	195086							
(lb MOLES/hr)	6382	7017							
(ACFM)	62800	49868							
(SCFM)	40316	44327							
LIQUID PHASE FLOW RATE									
(lb/hr)			2.20×10^6	2.19×10^6		11, 895	4149	415	3379
SOLID PHASE FLOW RATE									
(lb/hr)			245,241	245,241	300		461	415	46

LIMESTONE FGD PROCESS

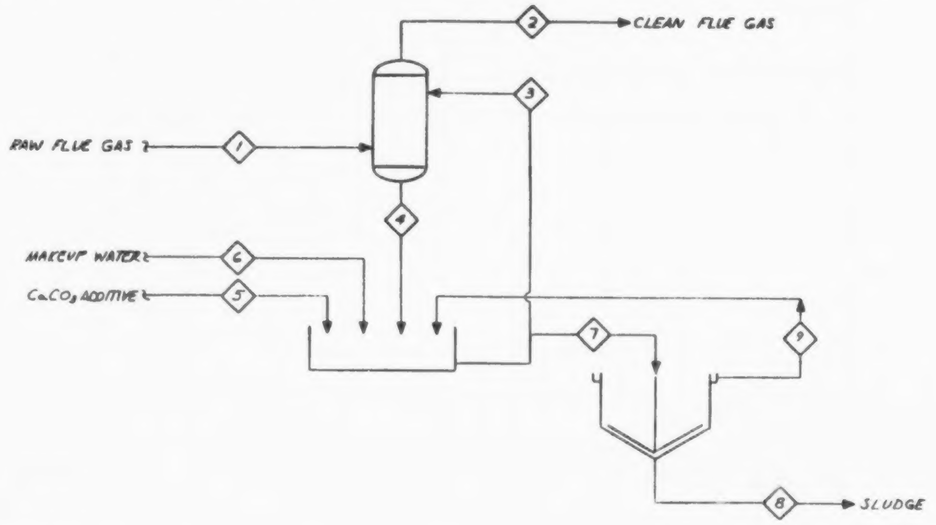
BOILER SIZE: 150×10^6 BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb/hr)	185092	195086	1.40×10^6	1.39×10^6	250	11,845	3840	692	2852
(GPM)	—	—	2535	2512		23.7	6.9	0.7	5.6
TEMPERATURE (°F)	350	125							
GAS FLOW RATE									
(lb/hr)	185092	195086							
(lb MOLES/hr)	6382	7017							
(ACFM)	62800	49868							
(SCFM)	40316	44327							
LIQUID PHASE FLOW RATE									
(lb/hr)			1.26×10^6	1.25×10^6		11,845	3457	346	2814
SOLID PHASE FLOW RATE									
(lb/hr)			140,780	140,780	250		384	346	38

LIMESTONE FGD PROCESS

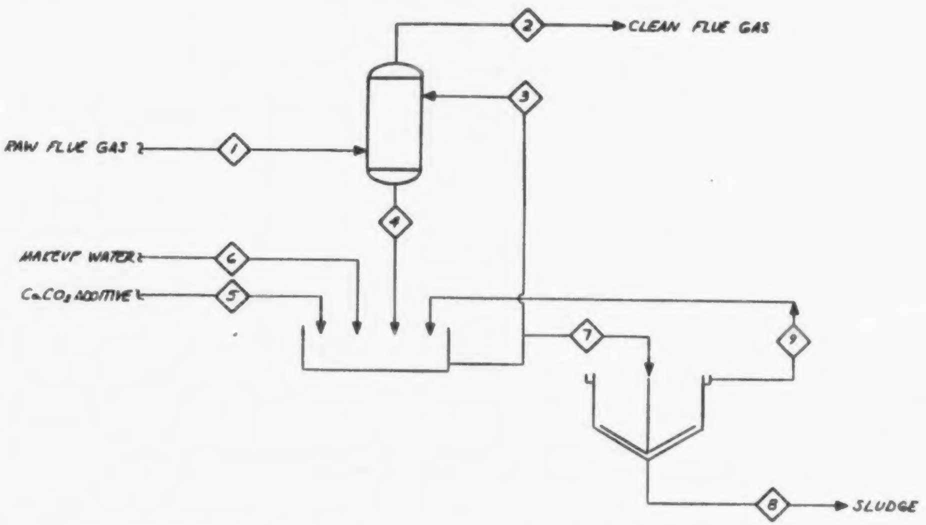
BOILER SIZE: 150 × 10⁶ BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb./hr.)	215,760	228,960	2.85×10^6	2.84×10^6	400	9196	6140	1106	4560
(GPM)	—	—	5139	5124		18.4	11.0	1.1	9.0
TEMPERATURE (°F)	350	125							
GAS FLOW RATE									
(lb./hr.)	215760	228,960							
(lb. MOLES/hr.)	7440	8148							
(ACFM)	73200	57,902							
(SCFM)	46993	51,468							
LIQUID PHASE FLOW RATE									
(lb./hr.)			2.56×10^6	2.55×10^6		9196	5526	553	4499
SOLID PHASE FLOW RATE									
(lb./hr.)			285,390	285,390	400		614	553	61

LIMESTONE FGD PROCESS

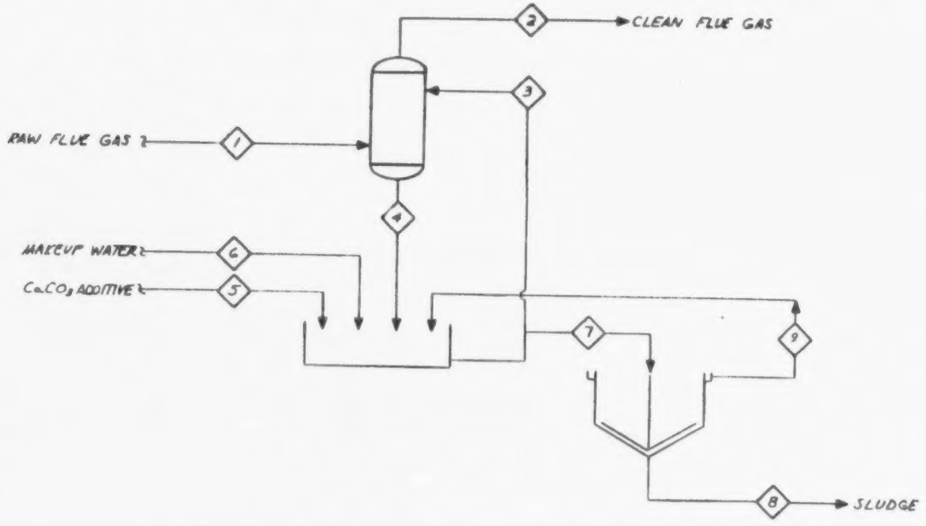
BOILER SIZE: 200 × 10^6 BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIER RETURN
TOTAL FLOW RATE									
(lb/hr)	215,760	278,960	2.24×10^6	7.24×10^6	378	9146	5800	1046	4307
(GPM)	—	—	4041	4026		18.3	10.4	1.0	8.5
TEMPERATURE (°F)	350	125							
GAS FLOW RATE									
(lb/hr)	215,760	228,960							
(lb. MOLES/hr)	7440	8148							
(ACFM)	73,200	57902							
(SCFM)	46,993	51,468							
LIQUID PHASE FLOW RATE									
(lb/hr)			2.01×10^6	2.00×10^6		9146	5220	523	4250
SOLID PHASE FLOW RATE									
(lb./hr)			224,455	224,455	378		580	523	57

LIMESTONE FGD PROCESS

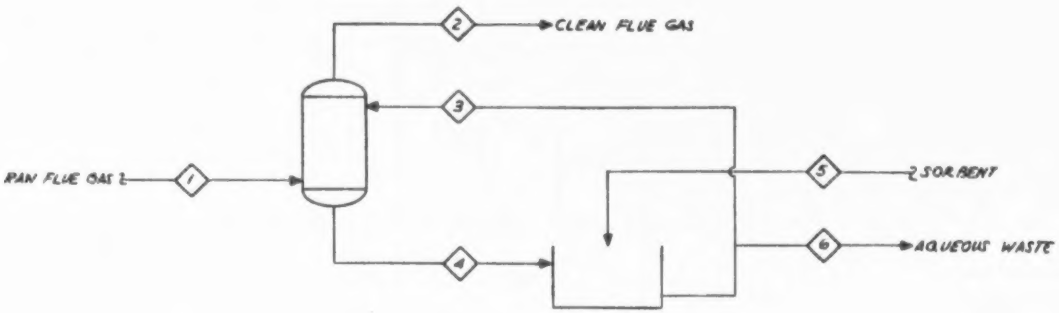
BOILER SIZE: 200×10^6 BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 85%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKEUP ALKALI	6 MAKEUP WATER	7 CLARIFIER FEED	8 SLUDGE	9 CLARIFIED RETURN
TOTAL FLOW RATE									
(lb/hr)	215760	228,960	1.63×10^6	1.62×10^6	333	9096	5110	920	3795
(GPM)	—	125	2945	2928	—	18.2	9.2	0.9	7.5
TEMPERATURE (°F)	350								
GAS FLOW RATE									
(lb/hr)	215760	228,960							
(lb MOLES/hr)	7440	8148							
(ACFM)	73200	57962							
(SCFM)	46993	51,468							
LIQUID PHASE FLOW RATE									
(lb/hr)			1.47×10^6	1.46×10^6		9096	4599	460	3744
SOLID PHASE FLOW RATE									
(lb./hr)			163560	163560	333		511	460	51

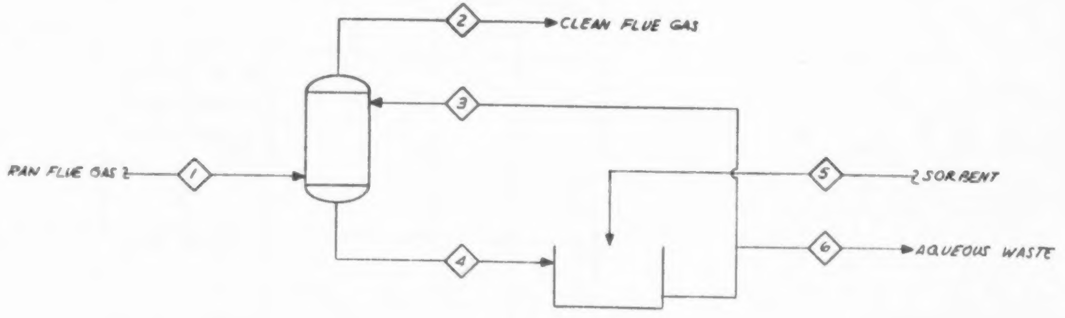
LIMESTONE FGD PROCESS

BOILER SIZE: 200 × 10^6 BTU/hr.
COAL TYPE: WESTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	35810	38336	70491	67842	9635	7040
(GPM)	—	—	134.3	129	18.7	13.4
TEMPERATURE (°F)	400	125	—	—	—	—
GAS FLOW RATE						
(lb/hr)	35810	38336				
(lb MOLES/hr.)	1234	1379				
(ACFM)	12900	9800				
(SCFM)	7800	8711				
LIQUID PHASE FLOW RATE						
(lb/hr.)			70491	67842	9635	7040
DISSOLVED SOLIDS WT %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb/hr.)						

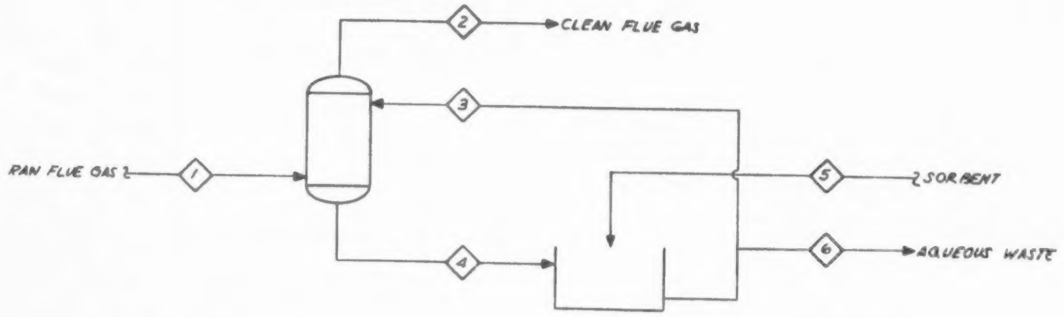
Sodium Throwaway FGD Process
BOILER SIZE: 30×10⁶ Btu/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	35810	38358	70528	67884	9162	6564
(GPM)	—	—	134.3	129	17.8	12.5
TEMPERATURE (°F)	400	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	35810	38358				
(lb. MOLES/hr.)	1235	1380				
(ACFM)	12900	9805				
(SCFM)	7800	8716				
LIQUID PHASE FLOW RATE						
(lb./hr.)			70528	67884	9162	6564
DISSOLVED SOLIDS WT. %						
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS

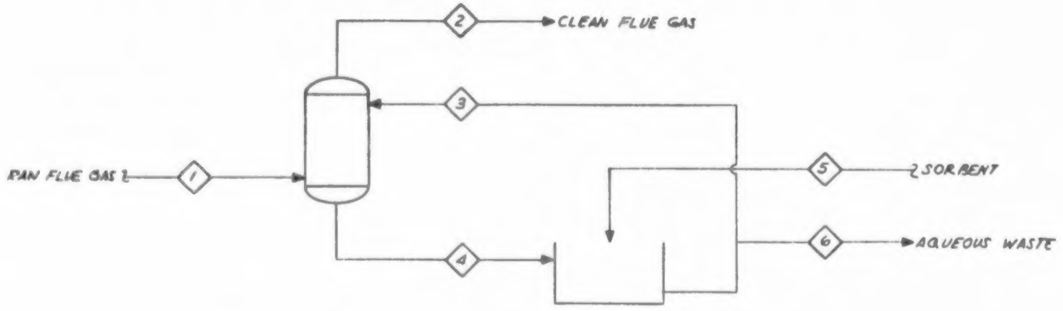
BOILER SIZE: 30×10⁶ BTU/HR
COAL TYPE: EASTERN
SO_2 REMOVAL: 85%



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	35810	38364	70529	67880	8484	5882
(GPM)	—	—	134.3	129	16.5	11.2
TEMPERATURE (°F)	400	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	35810	38364				
(lb. MOLES/hr.)	1234	1380				
(ACFM)	12900	9807				
(SCFM)	7800	8717				
LIQUID PHASE FLOW RATE						
(lb./hr.)			70529	67880	8484	5882
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

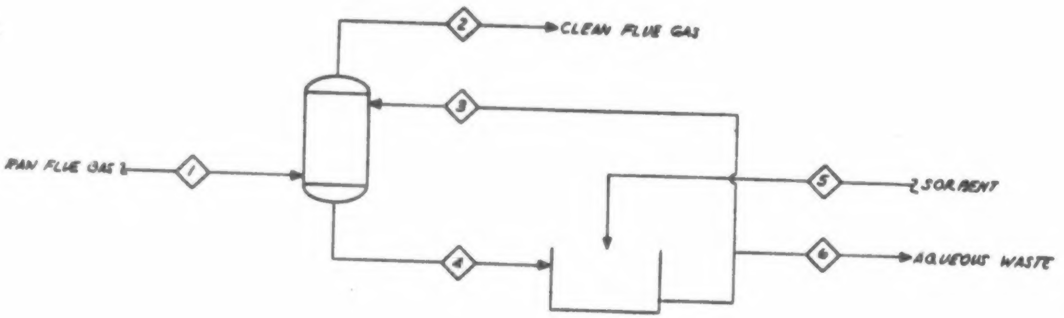
SODIUM THROWAWAY FGD PROCESS

BOILER SIZE: 30 X 10⁶ BTU/HR
COAL TYPE: EASTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	89664	96105	176593	169895	24726	18166
(GPM)	—	—	336.4	323	48	39.6
TEMPERATURE (°F)	400	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	89664	96105				
(lb. MOLES/hr.)	3092	3457				
(ACFM)	32300	24566				
(SCFM)	19530	21837				
LIQUID PHASE FLOW RATE						
(lb./hr.)			176593	169895	24726	18166
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

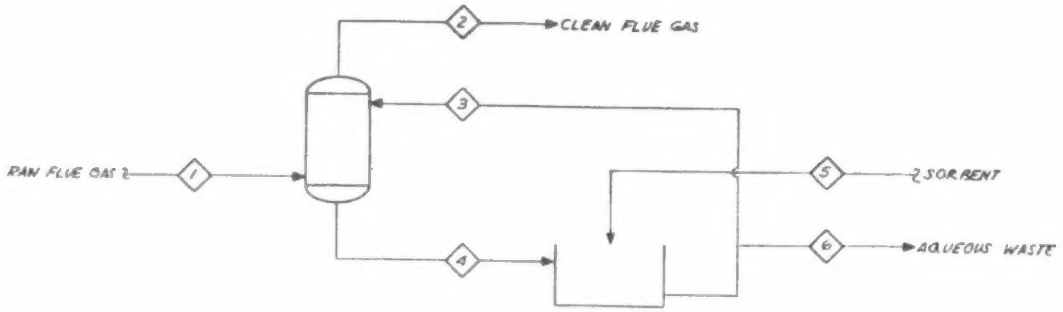
SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 75 x 10^6 BTU/HR
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



SODIUM THROWAWAY FGD PROCESS

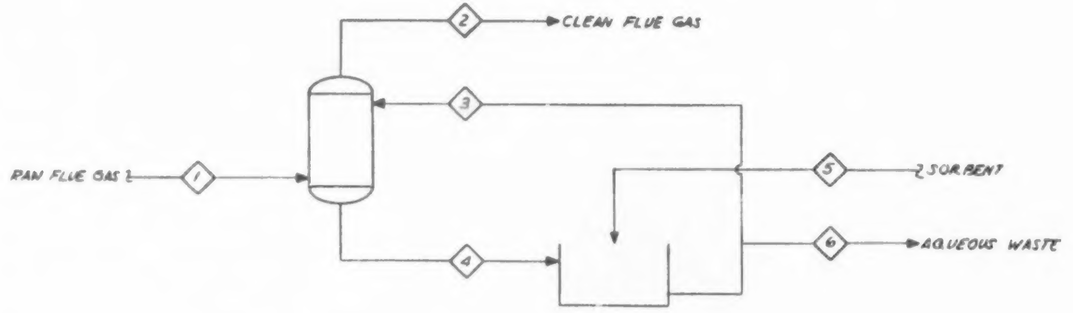
BOILER SIZE: 75×10^6 BTU/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 75%

STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	89664	96132	176646	169948	21338	14754
(GPM)	—	—	336.4	323	41.5	28.1
TEMPERATURE (°F)	400	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	89664	96132				
(lb. MOLES/hr.)	3092	3458				
(ACFM)	32300	24573				
(SCFM)	19530	21843				
LIQUID PHASE FLOW RATE						
(lb./hr.)			176646	169948	21338	14754
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						



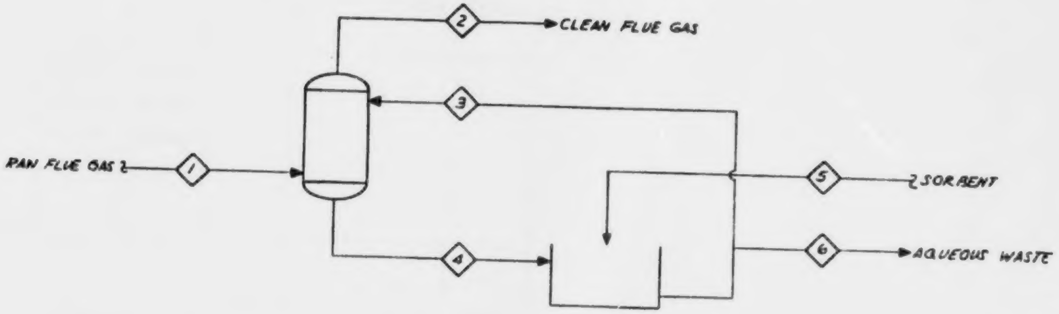
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	179883	192765	354300	340905	49554	36437
(GPM)	—	—	674.8	648	96.2	69.4
TEMPERATURE (°F)	400	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	179883	192765				
(lb. MOLES/hr.)	6203	6934				
(ACFM)	64800	49275				
(SCFM)	39181	43800				
LIQUID PHASE FLOW RATE						
(lb./hr.)			354300	340905	49554	36437
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 150 x10⁶ BTU/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 150 x 10^6 BTU/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 75%

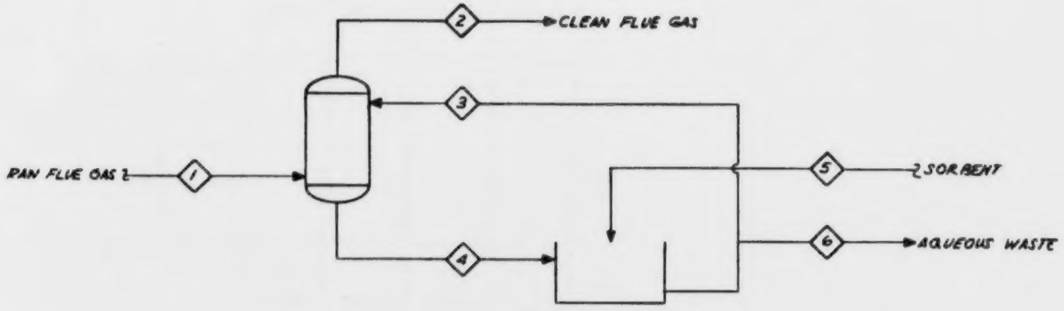
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	179883	192849	354327	340932	42520	29352
(GPM)	—	—	674.8	648	82.7	55.9
TEMPERATURE (°F)	400	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	179883	192849				
(lb. MOLES/hr.)	6203	6937				
(ACFM)	64800	49296				
(SCFM)	39181	43819				
LIQUID PHASE FLOW RATE						
(lb./hr.)			354327	340932	42520	29352
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	207643	222483	408997	393503	63434	48309
(GPM)	—	—	779	748	123	92
TEMPERATURE (°F)	900	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	207643	222483				
(lb. MOLES/hr.)	7160	8003				
(ACFM)	74800	56871				
(SCFM)	45228	50552				
LIQUID PHASE FLOW RATE						
(lb./hr.)			408997	393503	63434	48309
DISSOLVED SOLIDS WT. %			5	5	3.	5.
SOLID PHASE FLOW RATE						
(lb./hr.)						

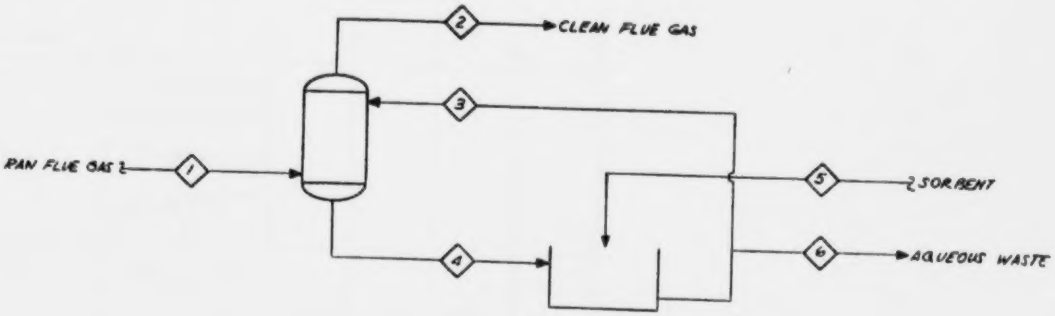
SODIUM Throwaway FGD PROCESS

BOILER SIZE: 200 x10⁶ Btu/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



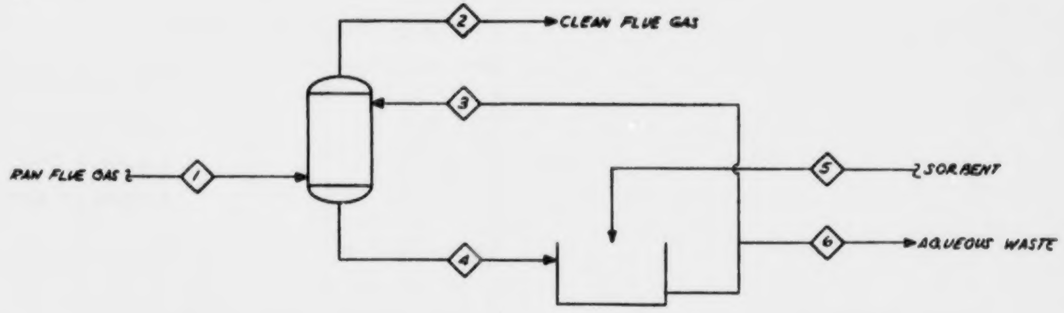
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	207643	222511	409106	393612	58949	43800
(GPM)	—	—	779	748	114.4	83.4
TEMPERATURE (°F)	400	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	207643	222511				
(lb. MOLES/hr.)	7160	8004				
(ACFM)	74800	56878				
(SCFM)	45228	50559				
LIQUID PHASE FLOW RATE						
(lb./hr.)			409106	393612	58949	43800
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 200X10⁶ Btu/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 85%



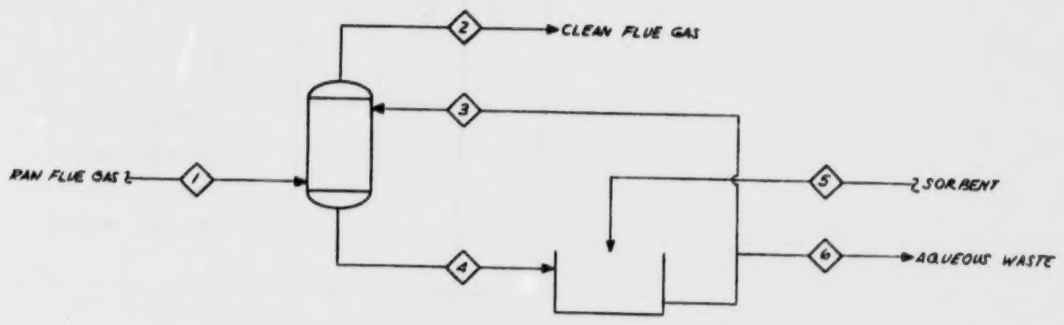
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	207643	222567	409105	393581	54415	39226
(GPM)	—	—	779	748	105.7	74.7
TEMPERATURE (°F)	900	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	207643	222567				
(lb. MOLES/hr.)	7160	8006				
(ACFM)	74800	56892				
(SCFM)	45228	50571				
LIQUID PHASE FLOW RATE						
(lb./hr.)			409105	393581	54415	39226
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 200 x 10^6 Btu/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 75%



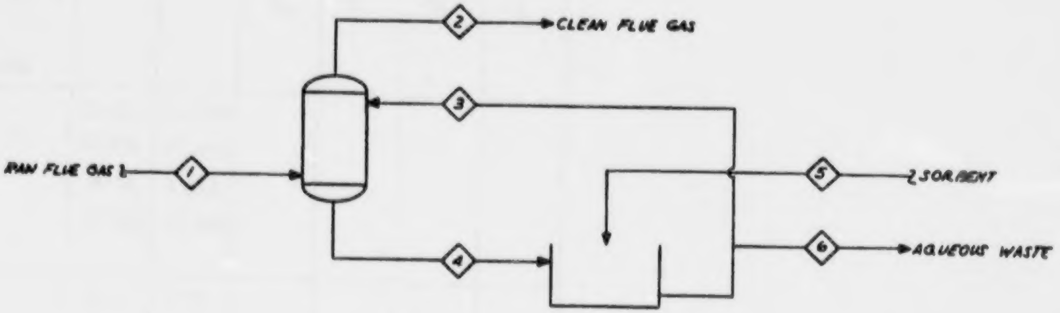
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	414,800	440,500	813,800	787,500	123,100	96,980
(GPM)	—	—	1550	1496	278	185
TEMPERATURE (°F)	400	130	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	414,800	440,500				
(lb. MOLES/hr.)	14,300	15,750				
(ACFM)	149,600	113,000				
(SCFM)	85,590	94,240				
LIQUID PHASE FLOW RATE						
(lb./hr.)						
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

Sodium Throwaway FGD Process
BOILER SIZE: 400 x10^6 Btu/hr
COAL TYPE: EASTERN
SO$_2$ REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	91,500	97,290	179,600	173,700	16,500	19,700
(GPM)	—	—	342	330	32	20
TEMPERATURE (°F)	400	130	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	91,500	97,290				
(lb. MOLES/hr.)	3155	3479				
(ACFM)	33,000	24,960				
(SCFM)	18,880	20,820				
LIQUID PHASE FLOW RATE						
(lb./hr.)						
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

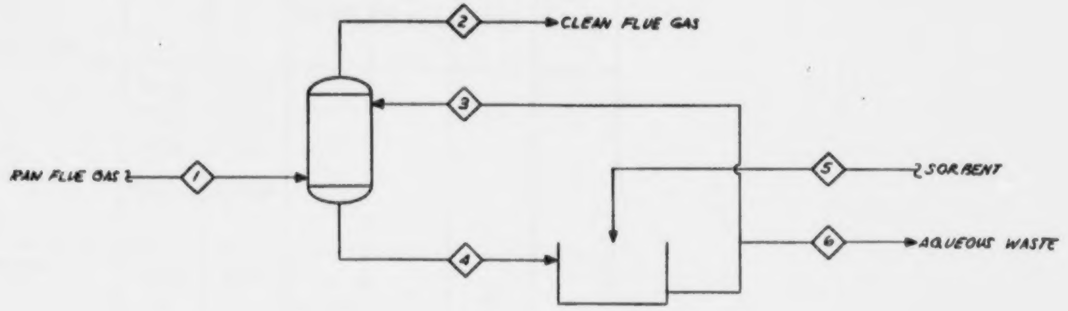
Sodium Throwaway FGD Process
BOILER SIZE: 75 x 10⁶ Btu/hr
COAL TYPE: medium Sulfur Coal
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	419,300	445,700	822,600	79[illegible]	83,270	56,970
(GPM)	—	—	1566	1512	162	108
TEMPERATURE (°F)	400	130	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	419,300	445,700				
(lb. MOLES/hr.)	14,460	15,940				
(ACFM)	151,200	114,400				
(SCFM)	86,500	95,370				
LIQUID PHASE FLOW RATE						
(lb./hr.)						
DISSOLVED SOLIDS WT. %			5	5	3	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS

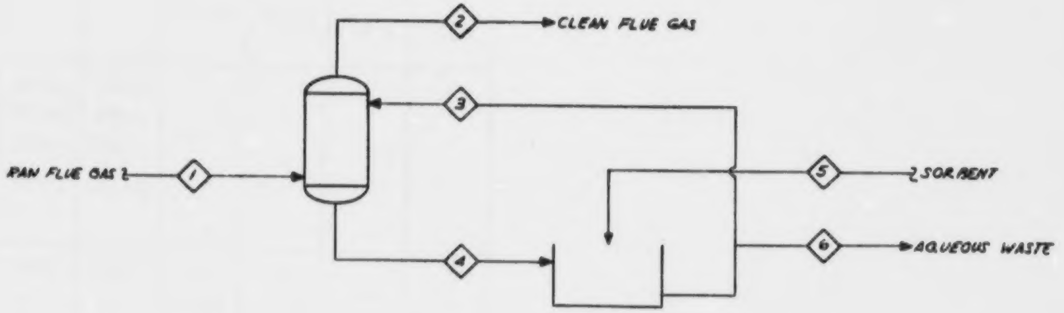
BOILER SIZE: 400 X10⁶ Btu/hr
COAL TYPE: MEDIUM Sulfur
SO_2 REMOVAL: 90 %



SODIUM THROWAWAY FGD PROCESS

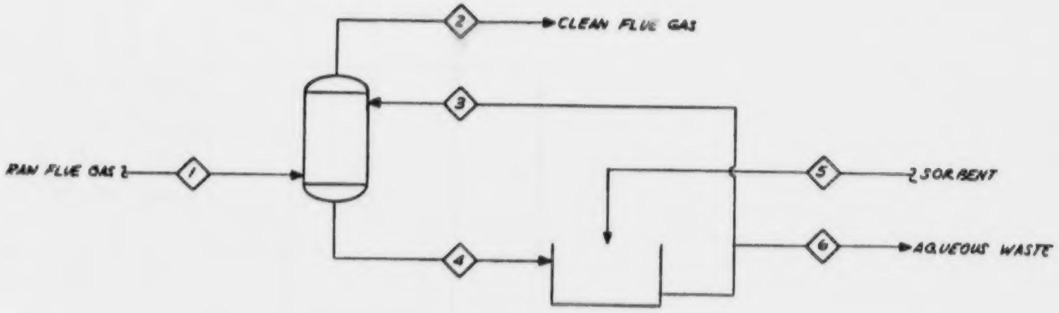
BOILER SIZE: 30×10^6 BTU/HR
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%

STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	36843	39116	68110	65811	3812	1525
(GPM)	—	—	129.6	125.0	7.5	2.9
TEMPERATURE (°F)	350	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	36843	39116				
(lb. MOLES/hr.)	1270	1397				
(ACFM)	12500	9927				
(SCFM)	8025	8824				
LIQUID PHASE FLOW RATE						
(lb./hr.)			68110	65811	3812	1525
DISSOLVED SOLIDS WT. %			5	5	2	5
SOLID PHASE FLOW RATE						
(lb./hr.)						



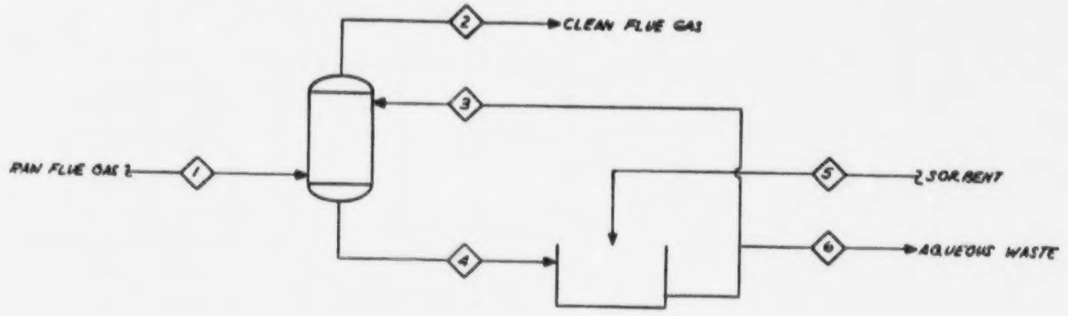
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	36843	39116	68146	65847	3655	1368
(GPM)	—	—	129.6	125	7.2	2.6
TEMPERATURE (°F)	350	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	36843	39116				
(lb. MOLES/hr.)	1270	1397				
(ACFM)	12500	9927				
(SCFM)	8026	8824				
LIQUID PHASE FLOW RATE						
(lb./hr.)			68146	65847	3655	1368
DISSOLVED SOLIDS WT. %			5	5	2	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

Sodium Throwaway FGD Process
BOILER SIZE: 30×10^6 BTU/HR
COAL TYPE: WESTERN
SO_2 REMOVAL: 85%



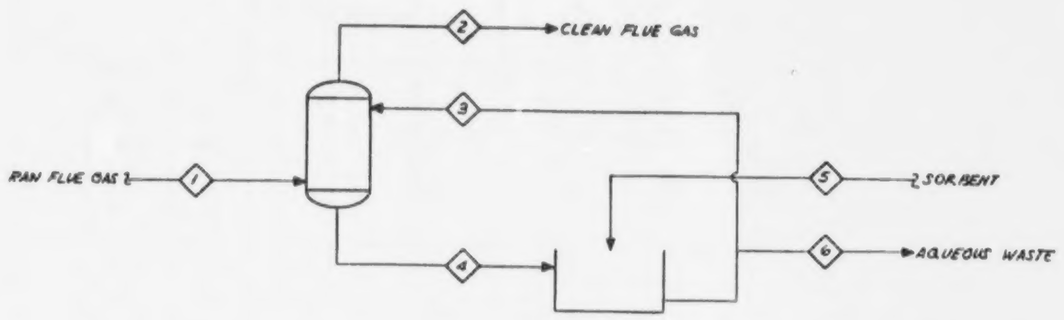
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	36843	39116	68133	65834	3550	1262
(GPM)	—	—	129.6	125.0	7.0	2.4
TEMPERATURE (°F)	350	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	36843	39116				
(lb. MOLES/hr.)	1270	1397				
(ACFM)	12500	9927				
(SCFM)	8025	8824				
LIQUID PHASE FLOW RATE						
(lb./hr.)			68133	65834	3550	1262
DISSOLVED SOLIDS WT. %			5	5	2	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 30×10⁶ BTU/HR
COAL TYPE: WESTERN
SO_2 REMOVAL: 75%



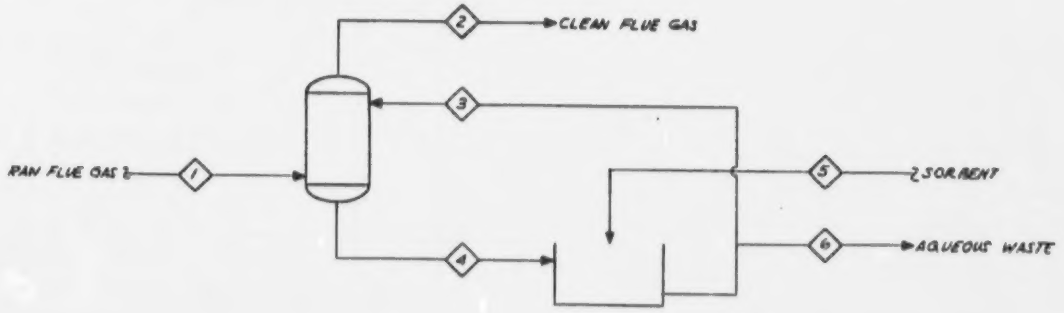
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	92539	98252	171076	165328	9401	3836
(GPM)	—	—	325.5	314	18.5	7.3
TEMPERATURE (°F)	350	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	92539	98252				
(lb. MOLES/hr.)	3191	3509				
(ACFM)	31400	24936				
(SCFM)	20158	22165				
LIQUID PHASE FLOW RATE						
(lb./hr.)			171076	165328	9401	3836
DISSOLVED SOLIDS WT. %			5	5	2	5.1
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 75x10⁶ Btu/hr
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	92539	98252	171122	165374	8822	3102
(GPM)	—	—	328.5	314	17.4	5.9
TEMPERATURE (°F)	350	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	92539	98252				
(lb. MOLES/hr.)	3191	3509				
(ACFM)	31400	24936				
(SCFM)	20158	22165				
LIQUID PHASE FLOW RATE						
(lb./hr.)			171122	165374	8822	3102
DISSOLVED SOLIDS WT. %			5	5	2	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

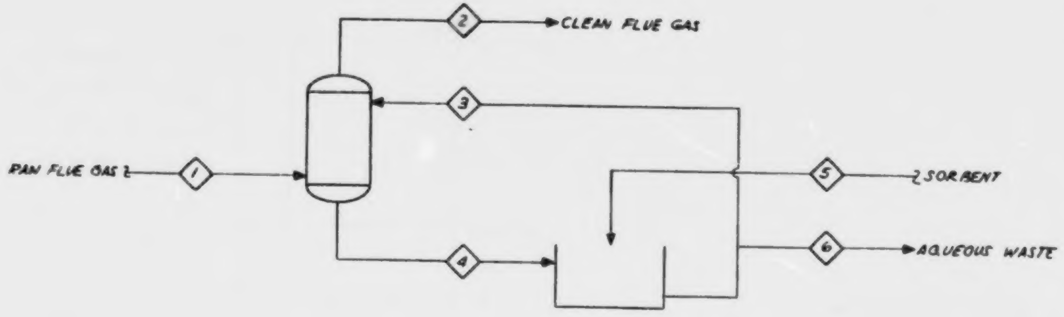
SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 75X10⁶ BTU/HR
COAL TYPE: WESTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	185092	197908	342051	330605	19052	7673
(GPM)	—	—	650.9	628	37.5	14.6
TEMPERATURE (°F)	350	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	185092	197908				
(lb. MOLES/hr.)	6382	7018				
(ACFM)	62800	49872				
(SCFM)	40316	44330				
LIQUID PHASE FLOW RATE						
(lb./hr.)			342051	330605	19052	7673
DISSOLVED SOLIDS WT %			5	5	2	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

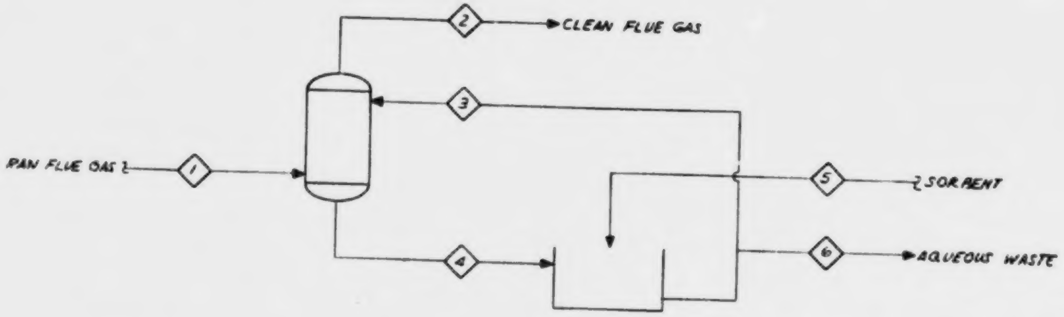
Sodium Throwaway FGD Process

BOILER SIZE: 150 x 10^6 BTU/hr
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



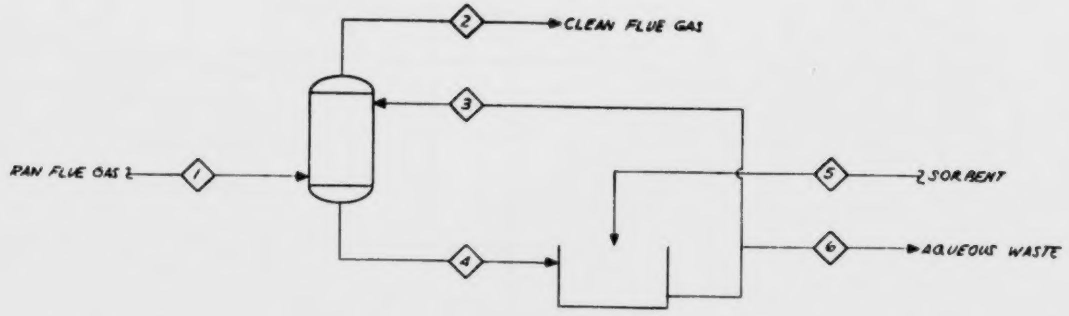
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	185092	197908	342168	330722	17593	6204
(GPM)	—	—	650.9	628	34.7	11.8
TEMPERATURE (°F)	350	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	185092	197908				
(lb. MOLES/hr.)	6382	7018				
(ACFM)	62800	49872				
(SCFM)	40316	44330				
LIQUID PHASE FLOW RATE						
(lb./hr.)			342168	330722	17593	6204
DISSOLVED SOLIDS WT. %			5	5	2	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS
DOILER SIZE: 150×10^6 BTU/hr
COAL TYPE: WESTERN
SO_2 REMOVAL: 75%



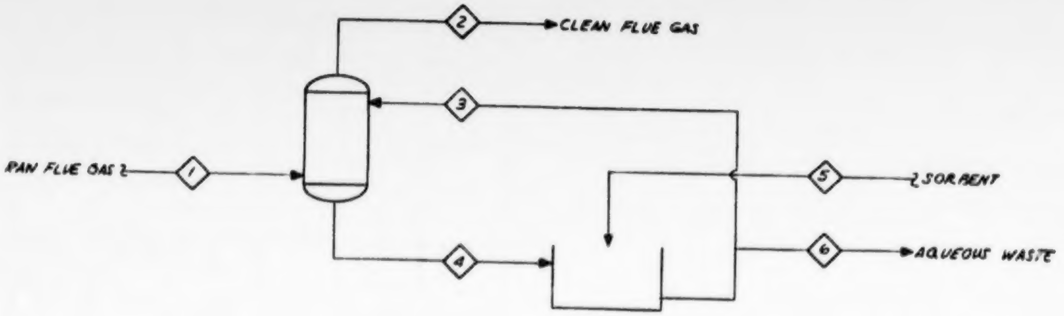
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	215744	229040	398780	385386	23554	10297
(GPM)	—	—	758.8	732	463	19.5
TEMPERATURE (°F)	350	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	215744	229040				
(lb. MOLES/hr.)	7439	8180				
(ACFM)	73200	58129				
(SCFM)	46993	51670				
LIQUID PHASE FLOW RATE						
(lb./hr.)			398780	385386	23554	10297
DISSOLVED SOLIDS WT. %			5	5	2	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 200×10^6 BTU/hr
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



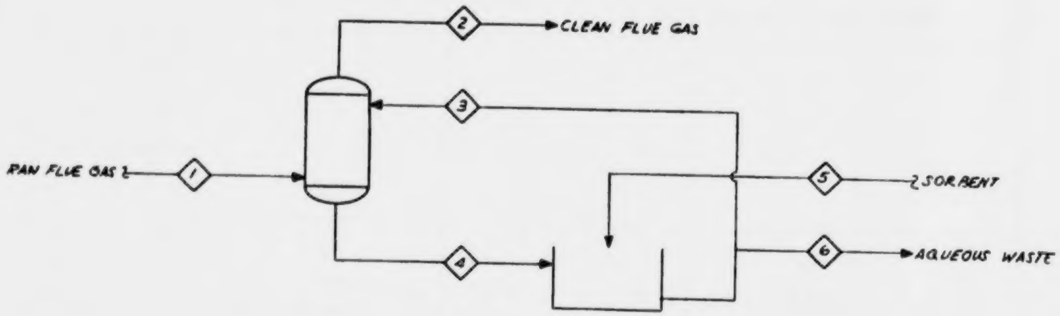
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	215744	229040	398917	385523	22564	9251
(GPM)	—	—	758.8	732	44.4	17.6
TEMPERATURE (°F)	900	125	—	—	—	—
GAS FLOW RATE						
(lb/hr)	215744	229040				
(lb MOLES/hr.)	7439	8180				
(ACFM)	73200	58129				
(SCFM)	46993	51670				
LIQUID PHASE FLOW RATE						
(lb/hr)			398917	385523	22564	9251
DISSOLVED SOLIDS WT. %			5	5	2	5
SOLID PHASE FLOW RATE						
(lb/hr)						

SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 200 x 10⁶ BTU/hr
COAL TYPE: WESTERN
SO_2 REMOVAL: 85%



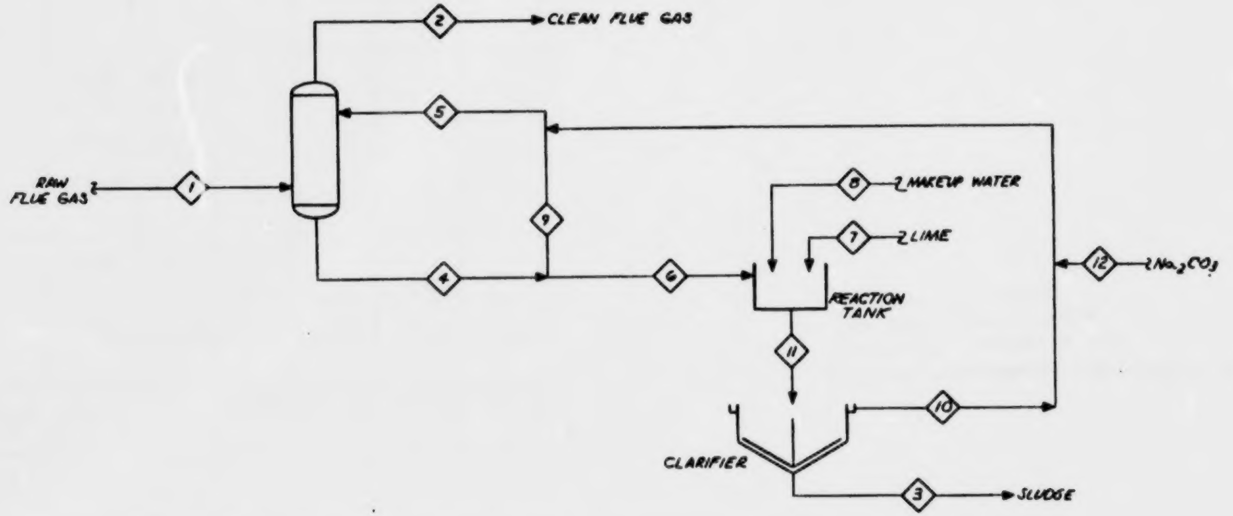
STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	215744	229040	398886	385492	21576	8254
(GPM)	—	—	758.8	732	42.5	15.7
TEMPERATURE (°F)	350	125	—	—	—	—
GAS FLOW RATE						
(lb./hr.)	215744	229090				
(lb. MOLES/hr.)	7439	8180				
(ACFM)	73200	58129				
(SCFM)	46993	51670				
LIQUID PHASE FLOW RATE						
(lb./hr.)			398886	385492	21576	8254
DISSOLVED SOLIDS WT. %			5	5	0	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 200 x 10^6 Btu/hr
COAL TYPE: WESTERN
SO_2 REMOVAL: 75%



STREAM NUMBER & DESCRIPTION	1 FLUE GAS	2 EXIT GAS	3 RECYCLE LIQUOR	4 SCRUBBER BOTTOMS	5 MAKE UP SORBENT	6 WASTE TO DISPOSAL
TOTAL FLOW RATE						
(lb./hr.)	429,800	454,800	793,600	768,600	45,500	20,500
(GPM)	—	—	1510	1460	89	39
TEMPERATURE (°F)	350	130	—	—		—
GAS FLOW RATE						
(lb./hr.)	429,800	454,800				
(lb. MOLES/hr.)	14,820	16,210				
(ACFM)	146,000	116,300				
(SCFM)	88,680	96,990				
LIQUID PHASE FLOW RATE						
(lb./hr.)						
DISSOLVED SOLIDS WT. %			5	5	2	5
SOLID PHASE FLOW RATE						
(lb./hr.)						

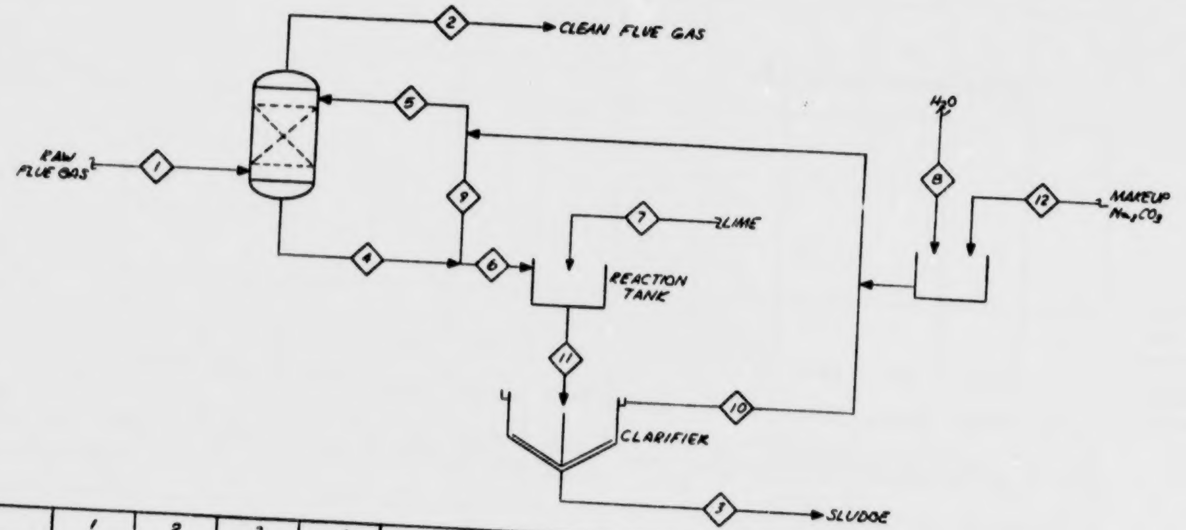
SODIUM THROWAWAY FGD PROCESS
BOILER SIZE: 400x10^6 Btu/hr
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1	2	3	4	5	6	7	8	9	10	11	12
	INLET GAS	EXIT GAS	SLUDGE	SCRUBBER BOTTOMS	SCRUBBER FEED	REGENER-ATION FEED	LIME	MAKE-UP H_2O	RECYCLE LIQUOR	CLARIFIER RETURN	CLARIFIER FEED	Na_2CO_3
TOTAL FLOW RATE												
(lb./hr.)	35810	38216	624	68125	70412	28302	131	2855	39823	30576	31200	13
(GPM)	—	—	0.6	129	134	53.6	—	5.7	75.4	58.7	59.3	—
TEMPERATURE (°F)	400	125	—	—	—	—	—	—	—	—	—	—
GAS FLOW RATE												
(lb./hr.)	35810	38216										
(lb moles/hr.)	1234	1374										
(ACFM)	12900	9764										
(SCFM)	7800	8679										
LIQUID PHASE FLOW RATE												
(lb./hr.)			312	68125	70412	28302		2855	39576	30576	30888	
DISSOLVED SOLIDS WT %			4	5	5	5			5	4	4	
SOLID PHASE FLOW RATE												
(lb./hr.)			312				131				312	13

DOUBLE ALKALI
FGD PROCESS

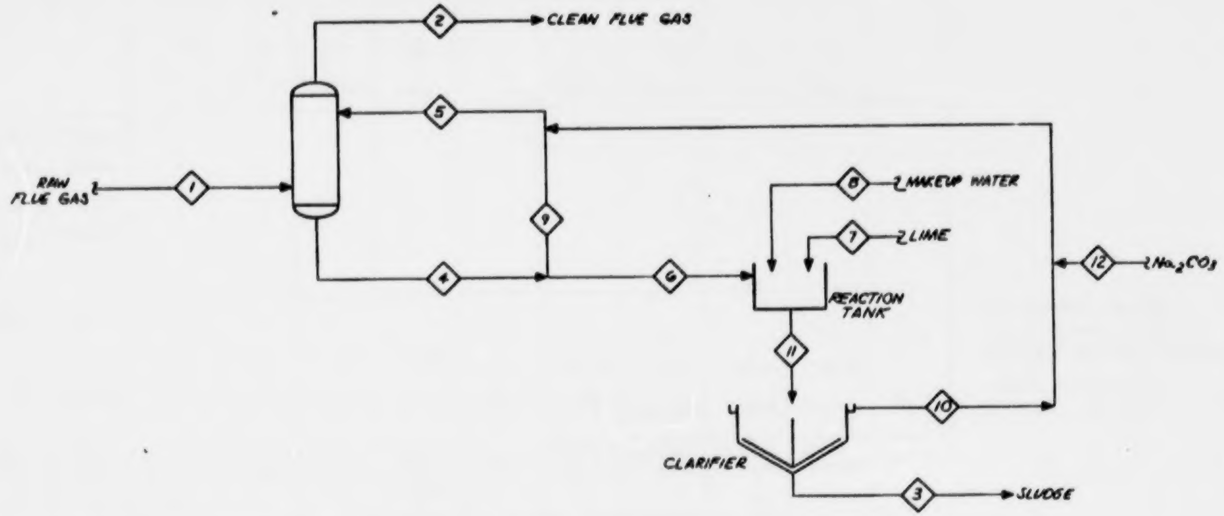
BOILER SIZE: 30×10^6 BTU/HR
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 SLUDGE	4 SCRUBBER BOTTOMS	5 SCRUBBER FEED	6 REGENERATION FEED	7 LIME	8 MAKE-UP H_2O	9 RECYCLE LIQUOR	10 CLARIFIER RETURN	11 CLARIFIER FEED	12 Na_2CO_3
TOTAL FLOW RATE												
(lb./hr.)	36840	39000	122	71371	73527	6041	25	2263	65330	5929	6050	5.3
(GPM)	—	—	0.1	125	129	10.6	-	4.5	114	10.5	10.6	—
TEMPERATURE (°F)	350	125	-	-	-	-	-	—	—	-	—	—
GAS FLOW RATE												
(lb./hr.)	36840	39000										
(lb. MOLES/hr.)	1270	1393										
(ACFM)	12500	9900										
(SCFM)	8025	8800										
LIQUID PHASE FLOW RATE												
(lb./hr.)			61	71371	73527	6041		2263	65330	5929	6050	
DISSOLVED SOLIDS WT. %			12	13	12	13		—	13	12	12	
SOLID PHASE FLOW RATE												
(lb./hr.)			61				25				61	5.3

DOUBLE ALKALI
FGD PROCESS

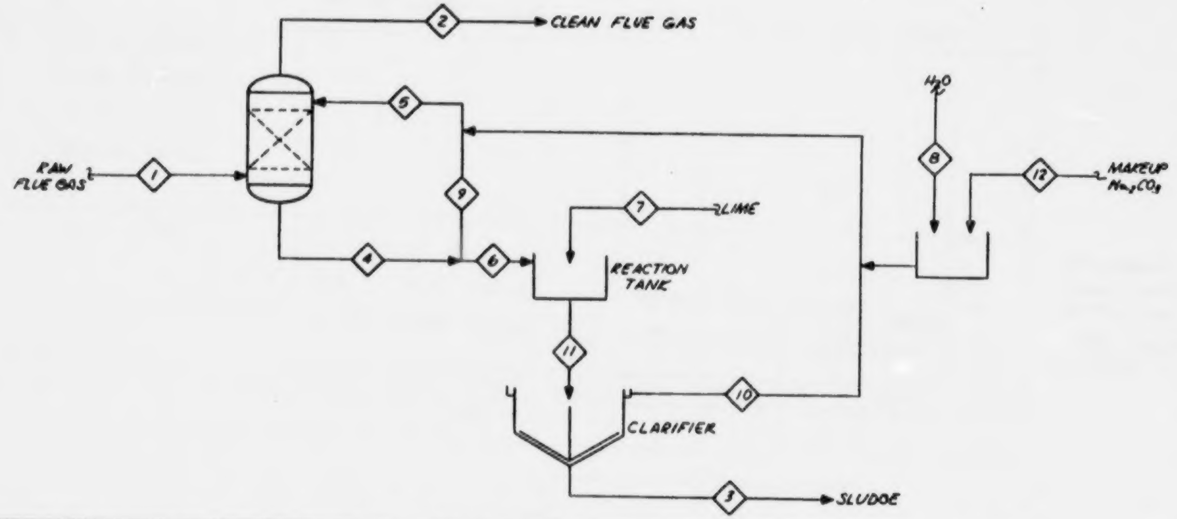
BOILER SIZE: 30×10^6 BTU/HR
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 SLUDGE	4 SCRUBBER BOTTOMS	5 SCRUBBER FEED	6 REGENERATION FEED	7 LIME	8 MAKE-UP H_2O	9 RECYCLE LIQUOR	10 CLARIFIER RETURN	11 CLARIFIER FEED	12 Na_2CO_3
TOTAL FLOW RATE												
(lb./hr.)	89660	96056	1584	170567	176320	71895	333	7196	98672	77616	79200	32
(GPM)	—	—	1.5	323	336	136	—	14.4	187	149	150	—
TEMPERATURE (°F)	900	125	—	—	—	—	—	—	—	—	—	—
GAS FLOW RATE												
(lb./hr.)	89660	96088										
(lb MOLES/hr.)	3092	3444										
(ACFM)	32300	24474										
(SCFM)	19530	21755										
LIQUID PHASE FLOW RATE												
(lb./hr.)			792	170567	176320	71895		7196	98672	77616	78408	
DISSOLVED SOLIDS WT %			4	5	5	5			5	4	4	
SOLID PHASE FLOW RATE												
(lb/hr)							333				792	32

DOUBLE ALKALI
FGD PROCESS

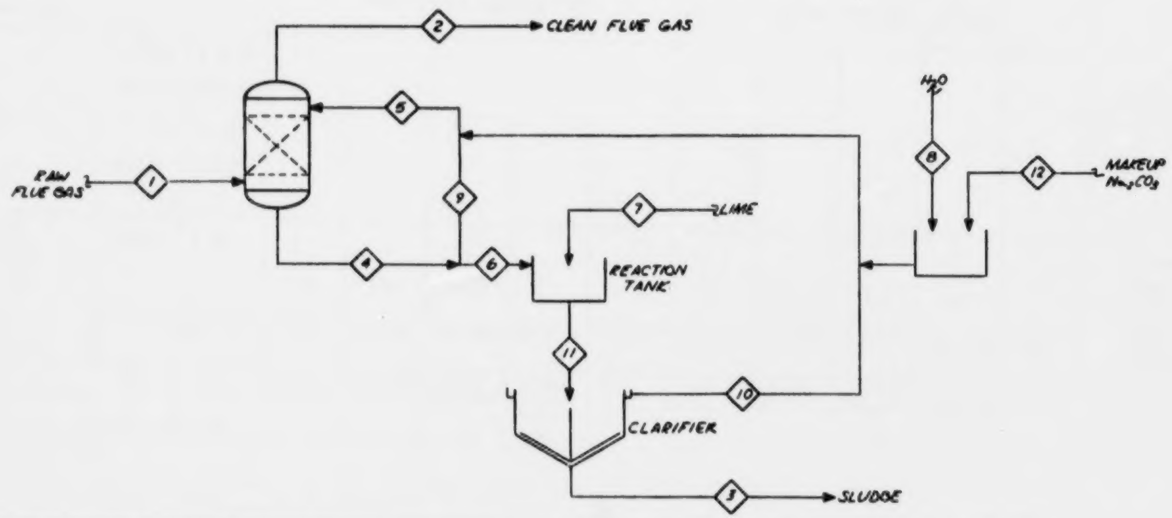
BOILER SIZE: 75×10^6 BTU/HR
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 SLUDGE	4 SCRUBBER BOTTOMS	5 SCRUBBER FEED	6 REGENERATION FEED	7 LIME	8 MAKE-UP H_2O	9 RECYCLE LIQUOR	10 CLARIFIER RETURN	11 CLARIFIER FEED	12 Na_2CO_3
TOTAL FLOW RATE												
(lb./hr)	91,500	97,320	888	177,726	183,230	38,026	186	6317	139,700	43,512	44,400	18
(GPM)	—	—	0.8	350	342	71	—	13	259	82	93	—
TEMPERATURE (°F)	400	130	—	—	—	—	—	—	—	—	—	—
GAS FLOW RATE												
(lb./hr)	91,500	97,320										
(lb. MOLES/hr)	3155	3480										
(ACFM)	33,000	24,970										
(SCFM)	18,880	20820										
LIQUID PHASE FLOW RATE												
(lb./hr)			444	177,726	183,230	38,026		6317	139,700	43,512	43,956	
DISSOLVED SOLIDS WT. %			5	7	7	7			7	5	5	
SOLID PHASE FLOW RATE												
(lb./hr)			444				186				444	18

DOUBLE ALKALI
FGD PROCESS

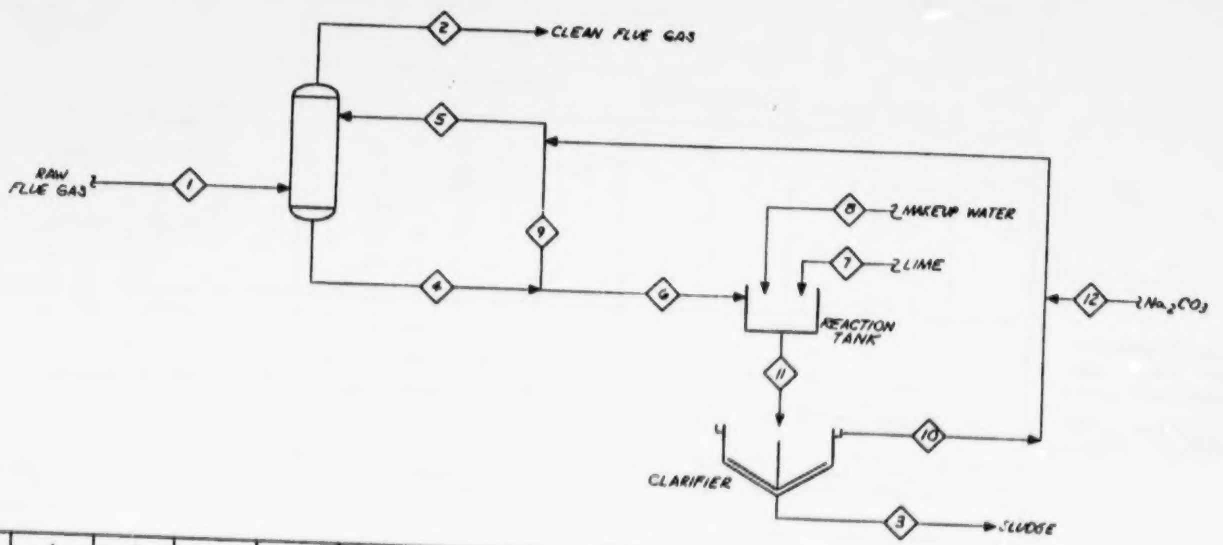
BOILER SIZE: 75×10^6 Btu/hr
COAL TYPE: MEDIUM S
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 SLUDGE	4 SCRUBBER BOTTOMS	5 SCRUBBER FEED	6 REGENERATION FEED	7 LIME	8 MAKE-UP H_2O	9 RECYCLE LIQUOR	10 CLARIFIER RETURN	11 CLARIFIER FEED	12 Na_2CO_3
TOTAL FLOW RATE												
(lb./hr.)	92546	98181	302	179152	184499	15078	63	5614	164074	14798	15100	13
(GPM)	—	—	0.3	314	325	26.4	—	11.2	288	26.2	26.4	—
TEMPERATURE (°F)	350	125	—	—	—	—	—	—	—	—	—	—
GAS FLOW RATE												
(lb./hr.)	92546	98181										
(lb. MOLES/hr.)	3191	3494										
(ACFM)	31400	24829										
(SCFM)	20158	22070										
LIQUID PHASE FLOW RATE												
(lb./hr.)			151	179152	184499	15078		5614	164074	14788	14949	
DISSOLVED SOLIDS WT. %			11	12	12	12			12	11	11	
SOLID PHASE FLOW RATE												
(lb./hr.)			151				63				151	13

DOUBLE ALKALI
FGD PROCESS

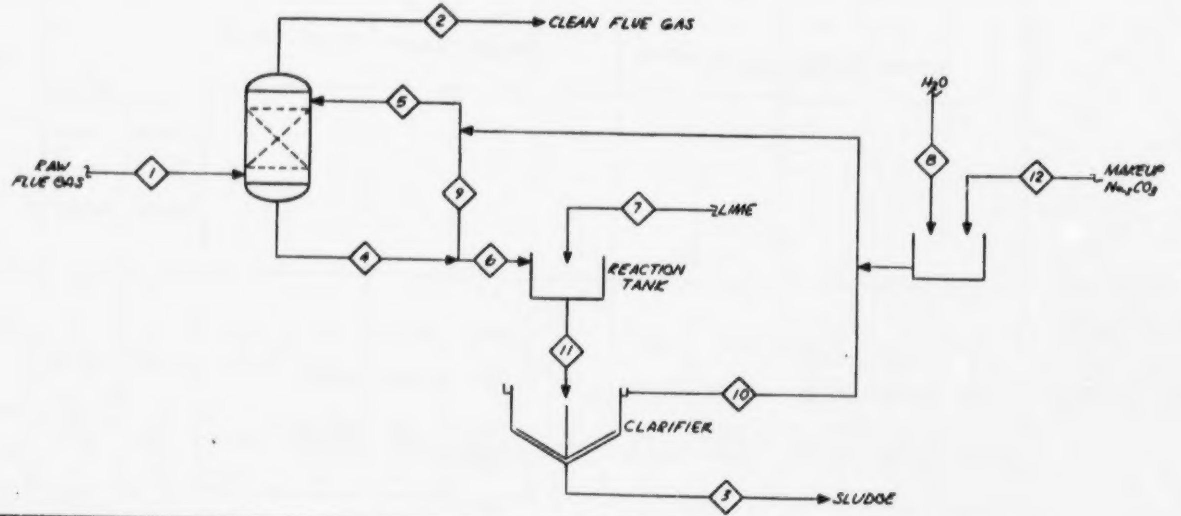
BOILER SIZE: 75x10⁶ BTU/HR
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 SLUDGE	4 SCRUBBER BOTTOMS	5 SCRUBBER FEED	6 REGENER-ATION FEED	7 LIME	8 MAKE-UP H_2O	9 RECYCLE LIQUOR	10 CLARIFIER RETURN	11 CLARIFIER FEED	12 Na_2CO_3
TOTAL FLOW RATE												
(lb./hr.)	207640	222920	4012	395191	408274	183587	885	16824	211604	196588	200600	82
(GPM)	—	—	3.8	798	777.8	347	–	33.7	400	377	381	–
TEMPERATURE (°F)	400	125	—	—	—	—	—	—	—	—	—	
GAS FLOW RATE												
(lb./hr.)	207640	222920										
(lb moles/hr.)	7160	7972										
(ACFM)	74800	56650										
(SCFM)	45228	50360										
LIQUID PHASE FLOW RATE												
(lb./hr.)			2006	395191	408274	183587		16824	211604	196585	200600	
DISSOLVED SOLIDS WT. %			4	5	5	5			5	4	4	
SOLID PHASE FLOW RATE												
(lb./hr.)												82

DOUBLE ALKALI
FGD PROCESS

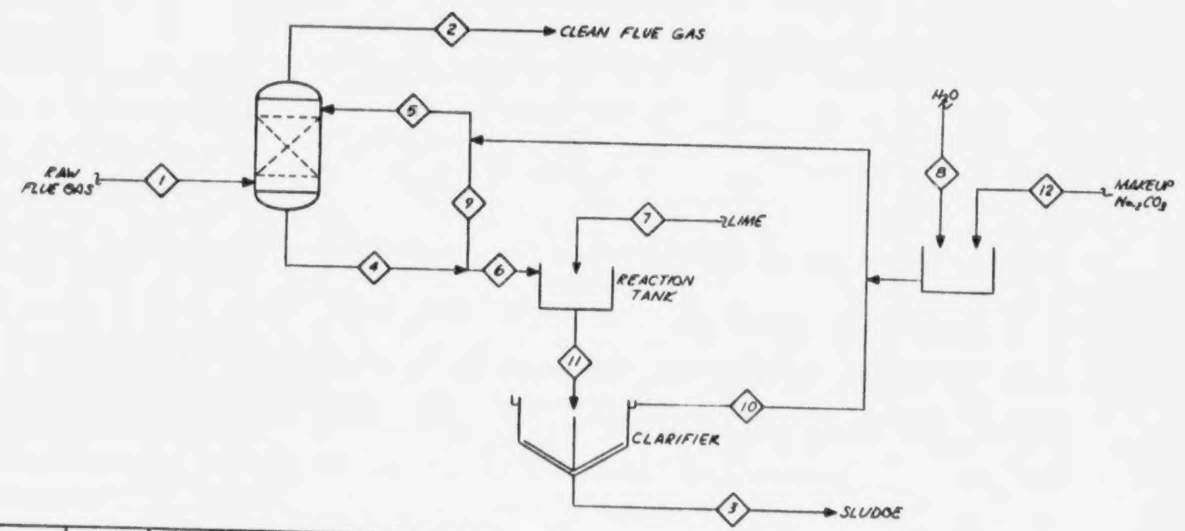
BOILER SIZE: 200×10^6 BTU/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 SLUDGE	4 SCRUBBER BOTTOMS	5 SCRUBBER FEED	6 REGENERATION FEED	7 LIME	8 MAKE-UP H_2O	9 RECYCLE LIQUOR	10 CLARIFIER RETURN	11 CLARIFIER FEED	12 Na_2CO_3
TOTAL FLOW RATE												
(lb./hr)	215760	228960	806	417764	429613	40242	168	13162	376922	39494	40300	35
(GPM)	—	—	0.7	732	758	70.6	—	26.3	661	69.9	70.6	—
TEMPERATURE (°F)	350	125	—	—	—	—	—	—	—	—	—	—
GAS FLOW RATE												
(lb/hr)	215760	228960										
(lb. MOLES/hr)	7440	8148										
(ACFM)	73200	57902										
(SCFM)	46993	51468										
LIQUID PHASE FLOW RATE												
(lb/hr)			403	417764	429613	40242		13162	376922	39494	40300	
DISSOLVED SOLIDS WT. %			11	12	12	12			12	11	11	
SOLID PHASE FLOW RATE												
(lb/hr)			403				168				403	35

DOUBLE ALKALI FGD PROCESS

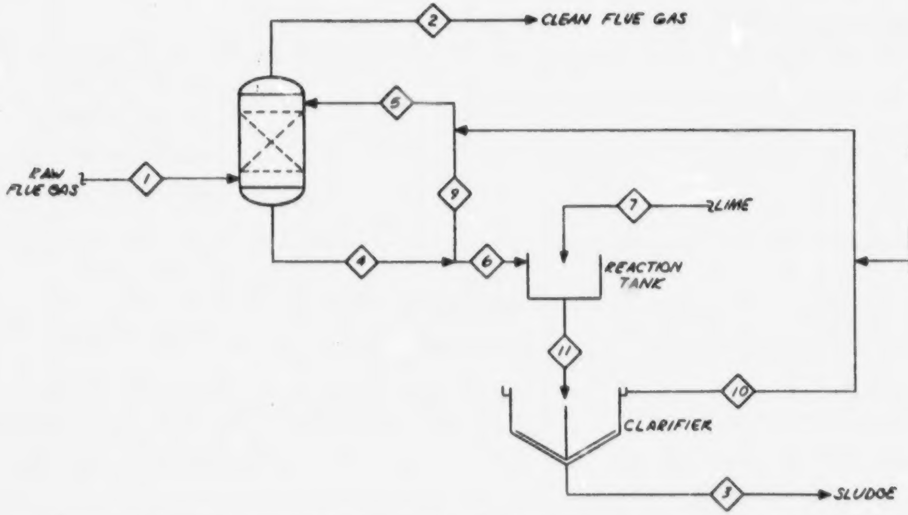
BOILER SIZE: 200×10^6 BTU/HR
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 SLUDGE	4 SCRUBBER BOTTOMS	5 SCRUBBER FEED	6 REGENERATION FEED	7 LIME	8 MAKE-UP H_2O	9 RECYCLE LIQUOR	10 CLARIFIER RETURN	11 CLARIFIER FEED	12 Na_2CO_3
TOTAL FLOW RATE												
(lb./hr.)	414,810	440,500	8450	789,682	812,767	391,138	1775	30,782	398,544	414,050	422,500	173
(GPM)	—	—	8.1	1496	1549	741	—	62	755	794	803	—
TEMPERATURE (°F)	400	130	—	—	—	—	—	—	—	—	—	—
GAS FLOW RATE												
(lb./hr.)	414,800	440,500										
(lb. MOLES/hr.)	14,305	15,753										
(ACFM)	149,600	113,030										
(SCFM)	85,590	94,255										
LIQUID PHASE FLOW RATE												
(lb./hr.)			4225	789,682	812,767	391,138		30,782	398,544	414,050	418,275	
DISSOLVED SOLIDS WT. %			4	5	5	5		—	5	4	4	
SOLID PHASE FLOW RATE												
(lb./hr.)			4225				1775					173

DOUBLE ALKALI
FGD PROCESS

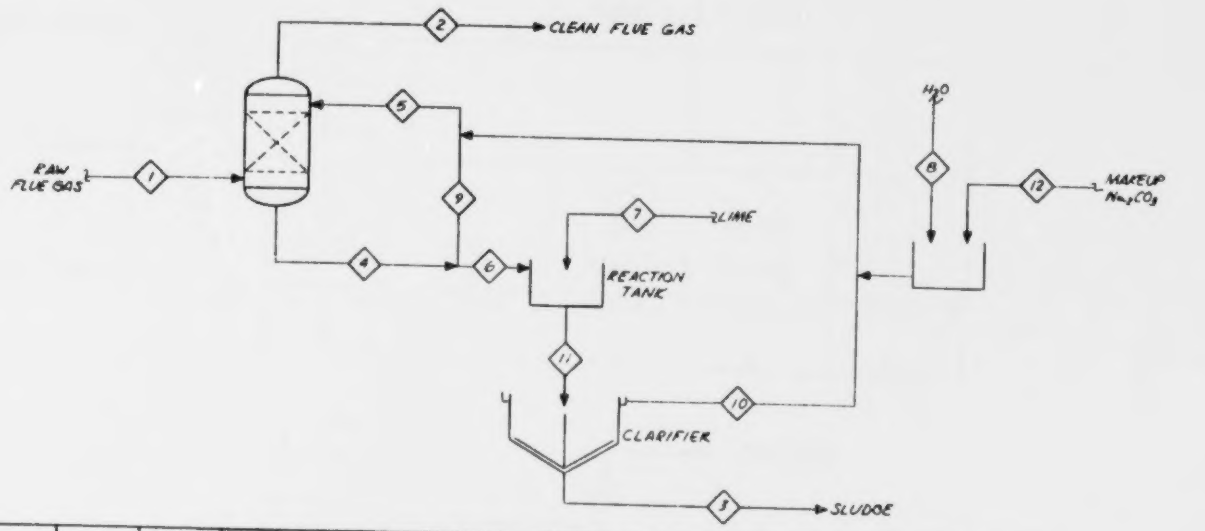
BOILER SIZE: 500×10^6 BTU/hr
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 SLUDGE	4 SCRUBBER BOTTOMS	5 SCRUBBER FEED	6 REGENERATION FEED	7 LIME	8 MAKE-UP H_2O	9 RECYCLE LIQUOR	10 CLARIFIER RETURN	11 CLARIFIER FEED	12 Na_2CO_3
TOTAL FLOW RATE												
(lb./hr.)	419,250	445,710	4732	814,332	839,293	207,006	990	29,260	607,326	231,868	236,600	99
(GPM)	—	—	4.5	1512	1566	384	—	59	1128	438	443	—
TEMPERATURE (°F)	900	130	—	—	—	—	—	—	—	—	—	—
GAS FLOW RATE												
(lb./hr.)	419,250	445,710										
(lb. MOLES/hr.)	14,457	15,939										
(ACFM)	151,200	114,360										
(SCFM)	86,500	95,369										
LIQUID PHASE FLOW RATE												
(lb./hr.)			2366	814,332	839,293	207,006		29,260	607,326	231,868	234,234	
DISSOLVED SOLIDS WT. %			6	7	7	7		—	7	6	6	
SOLID PHASE FLOW RATE												
(lb./hr.)							990				2366	99

DOUBLE ALKALI
FGD PROCESS

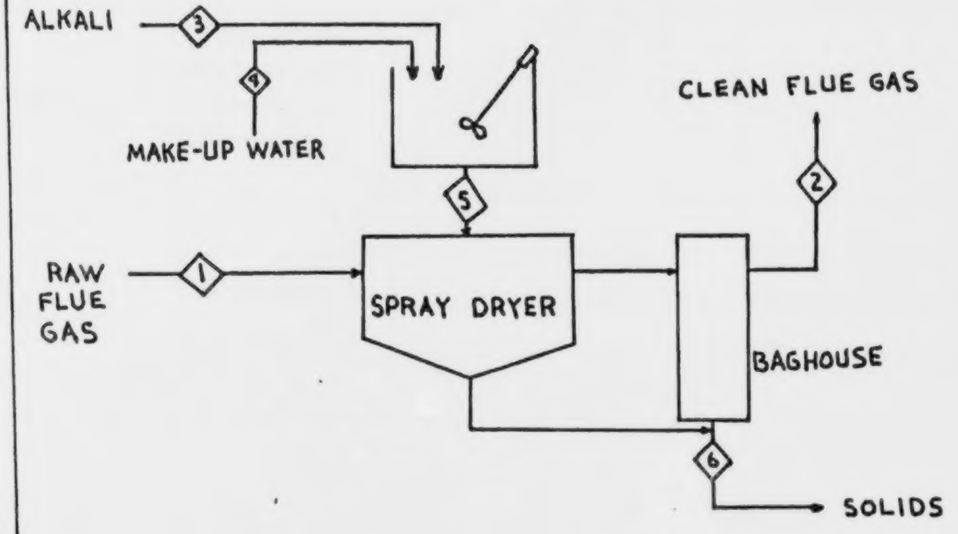
BOILER SIZE: 900×10^6 Btu/hr
COAL TYPE: MEDIUM Sulfur
SO_2 REMOVAL: 90%



STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 SLUDGE	4 SCRUBBER BOTTOMS	5 SCRUBBER FEED	6 REGENERATION FEED	7 LIME	8 MAKE-UP H_2O	9 RECYCLE LIQUOR	10 CLARIFIER RETURN	11 CLARIFIER FEED	12 Na_2CO_3
TOTAL FLOW RATE												
(lb./hr.)	429,800	459,800	1619	847,940	872,426	80,584	337	25,913	767,356	79,086	80,700	71
(GPM)	—	—	1.4	1460	1510	139	—	51.8	1321	137	139	—
TEMPERATURE (°F)	350	130	—	—	—	—	—	—	—	—	—	—
GAS FLOW RATE												
(lb./hr.)	429,800	459,800										
(lb. MOLES/hr.)	14,820	16,210										
(ACFM)	146,000	116,300										
(SCFM)	88,680	96,990										
LIQUID PHASE FLOW RATE												
(lb./hr.)			807	847,940	872,426	80,584		25,913	767,356	79,086	79,893	
DISSOLVED SOLIDS WT. %			13	14	14	14		—	14	13	13	
SOLID PHASE FLOW RATE												
(lb./hr.)			807				337				807	71

DOUBLE ALKALI
FGD PROCESS

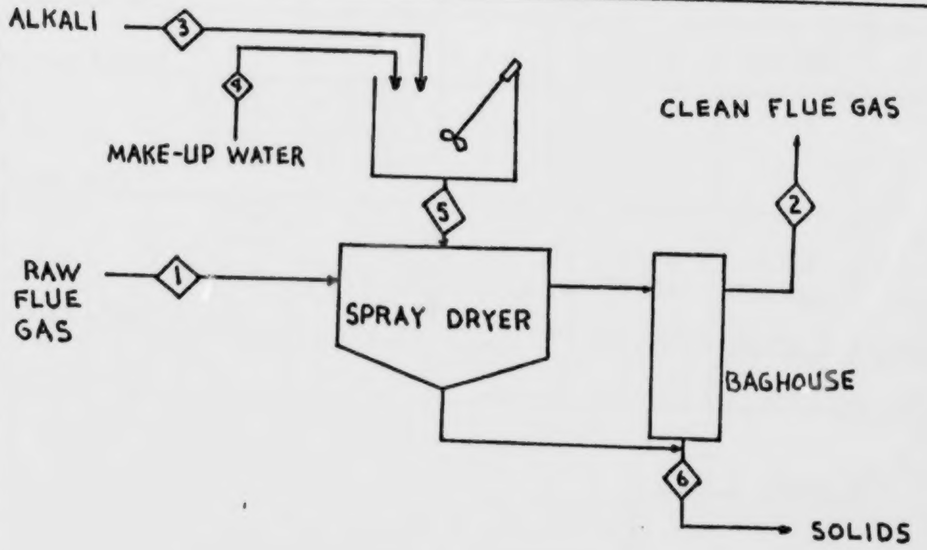
BOILER SIZE: 400x10^6 BTU/hr
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%



SPRAY DRYER FGD

Boiler Size: 75×10^6 Btu/hr
Coal Type: MEDIUM Sulfur
SO_2 Removal: 70%
Type Alkali: LIME

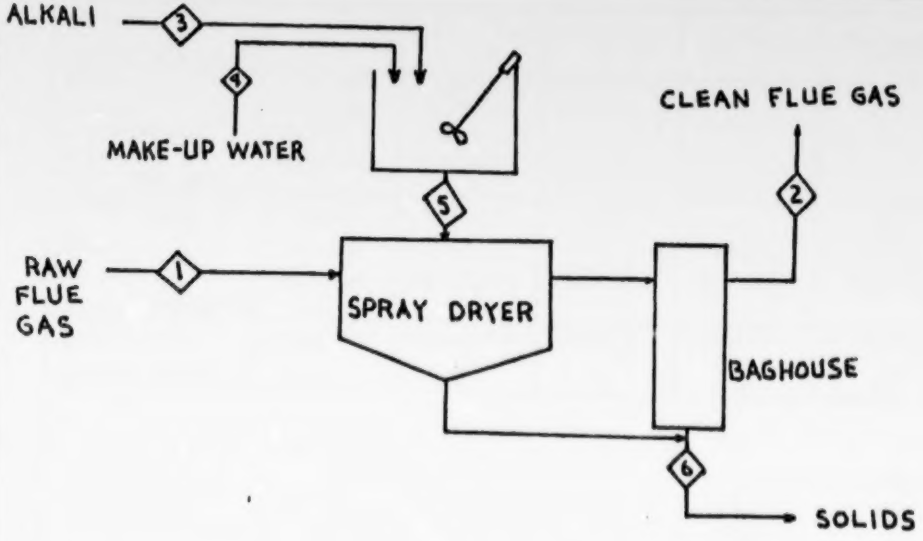
Stream Number & Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE lb/hr	91,691	96,359	261	4948	5209	675
gpm	—	—		9.9	9.9	
Temp (°F)	400	175				
GAS FLOW RATE lb/hr	91,503	96,358				
lb/moles/hr	3155	3427				
ACFM	33,000	26,467				
SCFM	18,879	20,507				
LIQUID FLOW RATE lb/hr				4948	4948	
SOLID FLOW RATE lb/hr						



SPRAY DRYER FGD

Boiler Size: 962×10^6 Btu/hr
Cool Type: Med. Sulfur
SO_2 Removal: 70%
Type Alkali: LIME

Stream Number & Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE						
lb/hr	422,488	441,532	1389	22,690	24,079	5782
gpm	—	—	—	45.4	45.4	—
Temp (°F)	400	175	—	—	—	—
GAS FLOW RATE						
lb/hr	419,250	441,516				
lb/moles/hr	14,457	15,703				
ACFM	151,200	121,270				
SCFM	86,500	93,959				
LIQUID FLOW RATE						
lb/hr				22,690	22,690	
SOLID FLOW RATE						
lb/hr	3198	16	1389		1389	5782



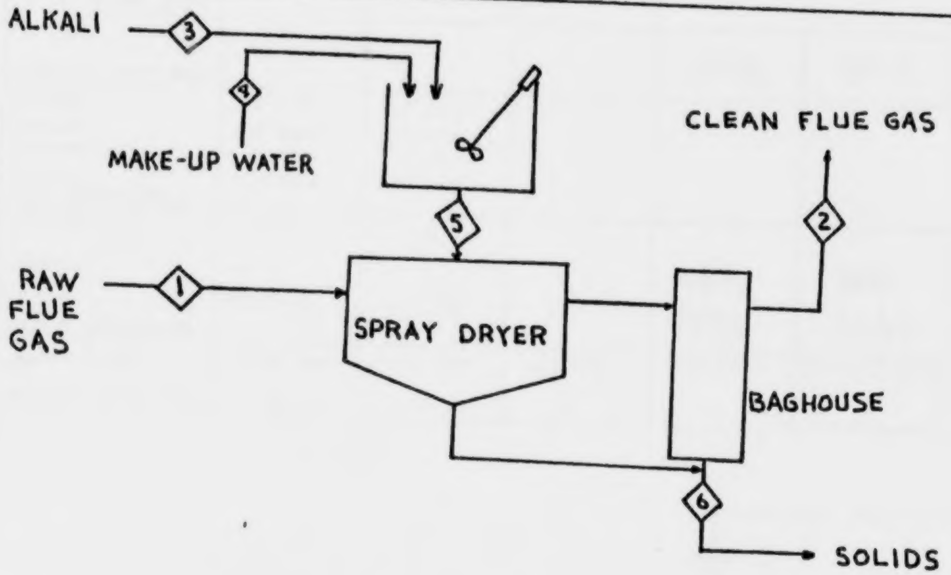
SPRAY DRYER FGD

Boiler Size: 400×10^6 Btu/hr
Cool Type: WESTERN
SO_2 Removal: 70%
Type Alkali: LIME

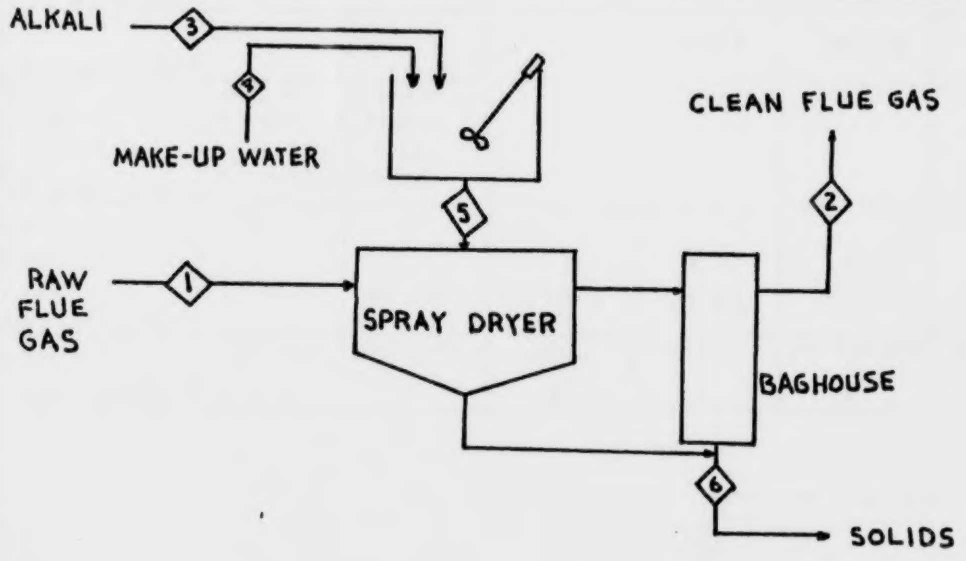
Stream Number / Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE						
lb/hr	431,620	451,532	499	21,890	22,389	2781
gpm	—	—	—	43.8	43.8	—
Temp (°F)	350	175	—	—	—	—
GAS FLOW RATE						
lb/hr	429,820	451,523				
lb/moles/hr	14,820	16,031				
ACFM	146,000	123,800				
SCFM	88,680	95,918				
LIQUID FLOW RATE						
lb/hr				21,890	21,890	
SOLID FLOW RATE						
lb/hr	1800	9	499		499	2781

SPRAY DRYER FGD

Boiler Size : 150×10^6 Btu/hr
Cool Type : Western
SO_2 Removal : 90 %
Type Alkali : Sodium



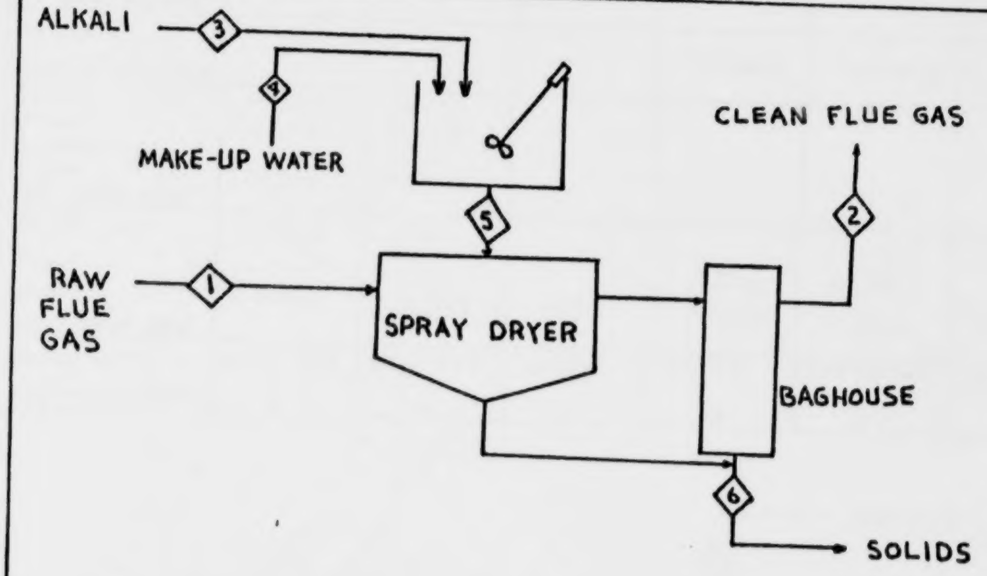
Stream Number & Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE						
lb/hr	185,640	190,794	428	9688	10116	1047
gpm				19.4	19.4	
Temp (°F)	350	175				
GAS FLOW RATE						
lb/hr	185,092	190,791				
lb/moles/hr	6382	6718				
ACFM	62,800	51,820				
SCFM	40,316	42,435				
LIQUID FLOW RATE						
lb/hr				9688	9688	
SOLID FLOW RATE						
lb/hr	548	3	428		428	1047



SPRAY DRYER FGD

Boiler Size: 150 x 10^6 Btu/hr
Coal Type: WESTERN
SO_2 Removal: 75%
Type Alkali: SODIUM

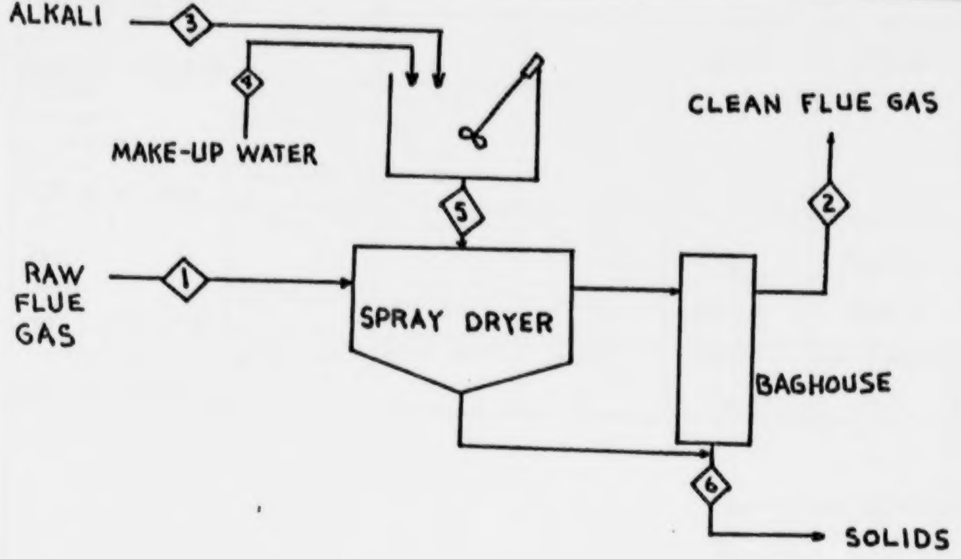
Stream Number & Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE						
lb/hr	185,640	190,735	311	9688	9999	904
gpm				19.4	19.4	
Temp (°F)	350	175				
GAS FLOW RATE						
lb/hr	185,092	190,732				
lb/moles/hr	6382	6716				
ACFM	62,800	51,805				
SCFM	40,316	42,423				
LIQUID FLOW RATE						
lb/hr				9688	9688	
SOLID FLOW RATE						
lb/hr	548	3	311		311	904



SPRAY DRYER FGD

Boiler Size: 150 x 10^6 BTU/HR
Coal Type: WESTERN
SO_2 Removal: 50 %
Type Alkali: SODIUM

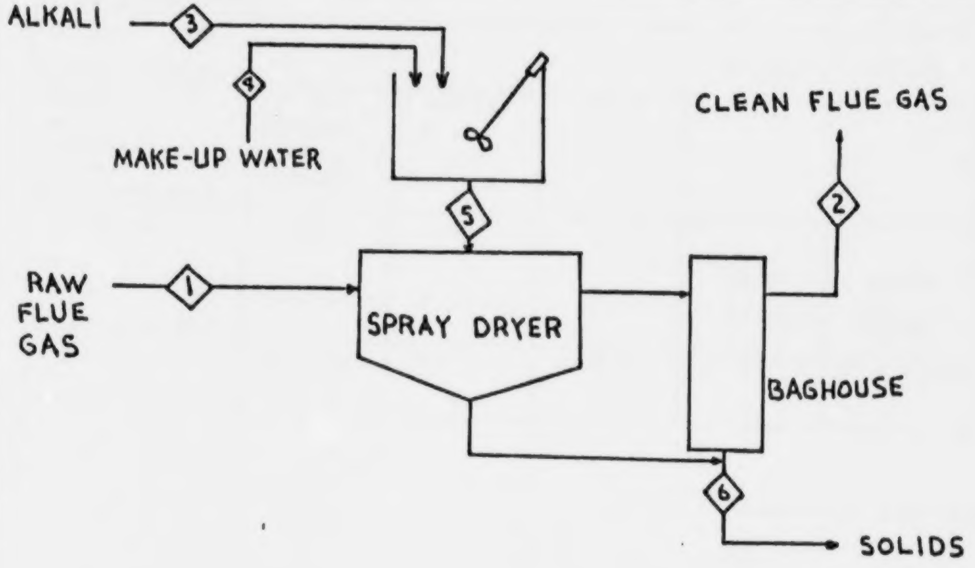
Stream Number & Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE						
lb/hr	185,640	190,794	214	9688	9902	785
gpm				19.4	19.4	
Temp (°F)	350	175				
GAS FLOW RATE						
lb/hr	185,092	190,791				
lb/moles/hr	6382	6718				
ACFM	62,800	51,820				
SCFM	40,316	42,435				
LIQUID FLOW RATE						
lb/hr				9688	9688	
SOLID FLOW RATE						
lb/hr	548	3	214		214	785



SPRAY DRYER FGD

Boiler Size: 200 x 10^6 Btu/hr
Coal Type: WESTERN
SO_2 Removal: 75 %
Type Alkali: SODIUM

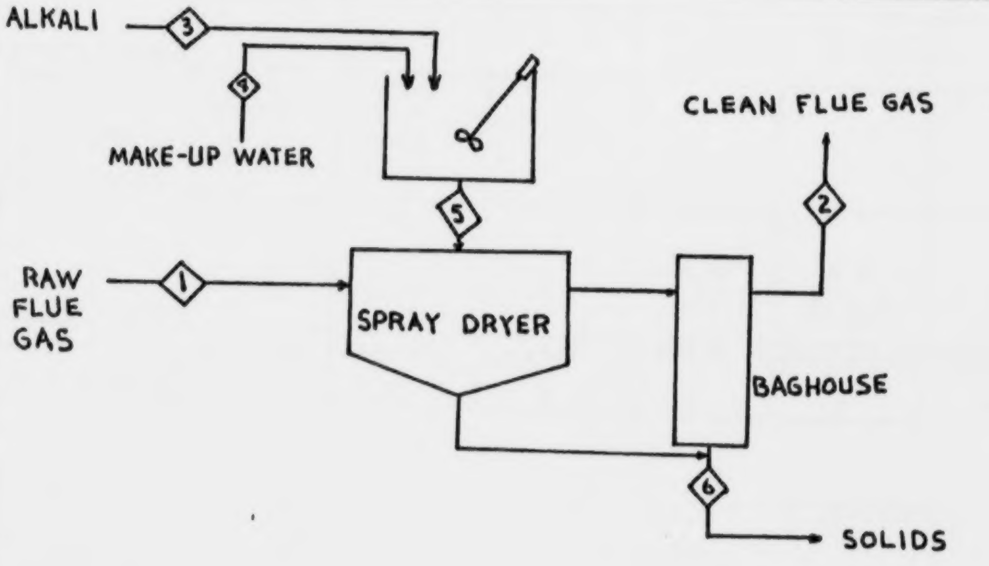
Stream Number & Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE						
lb/hr	216,644	222,405	415	11448	11863	1378
gpm				22.9	22.9	
Temp (°F)	350	175				
GAS FLOW RATE						
lb/hr	215,744	222,400				
lb/moles/hr	7439	7831				
ACFM	73,200	60,405				
SCFM	46,993	49,466				
LIQUID FLOW RATE						
lb/hr				11448	11448	
SOLID FLOW RATE						
lb/hr	900	5	415		415	1378



SPRAY DRYER FGD

Boiler Size: 60 x10^6 Btu/hr
Coal Type: Western
SO$_2$ Removal: 75%
Type Alkali: Sodium

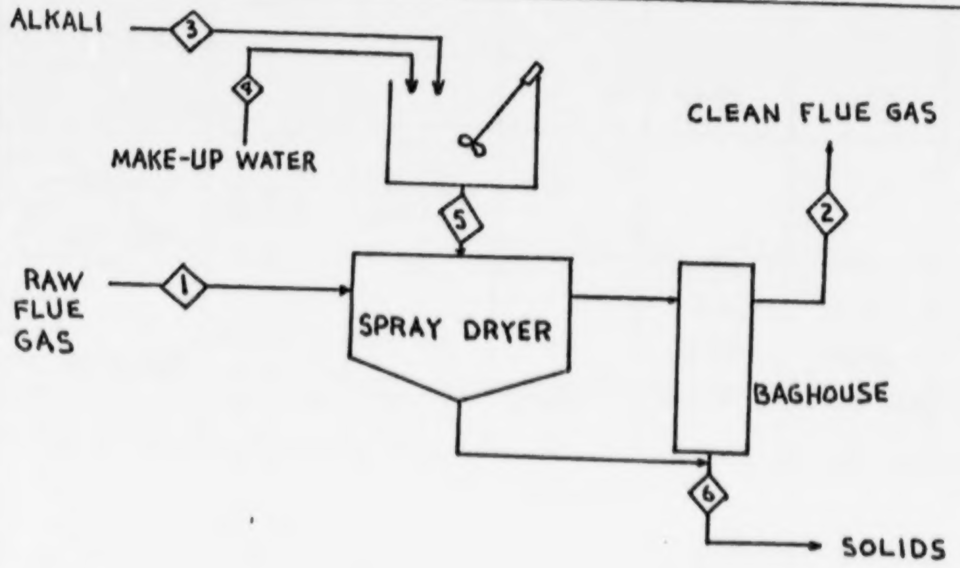
Stream Number & Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE						
lb/hr	73,767	75,970	123	3963	4086	227
gpm				7.9	7.9	
Temp (°F)	350	175				
GAS FLOW RATE						
lb/hr	73,683	75,970				
lb/moles/hr	2541	2675				
ACFM	25,000	20,634				
SCFM	16.049	16,897				
LIQUID FLOW RATE						
lb/hr				3963	3963	
SOLID FLOW RATE						
lb/hr	84	0.4	123		123	227



SPRAY DRYER FGD

Boiler Size: 150 x 10^6 Btu/hr
Coal Type: Western
SO_2 Removal: 90
Type Alkali: Lime

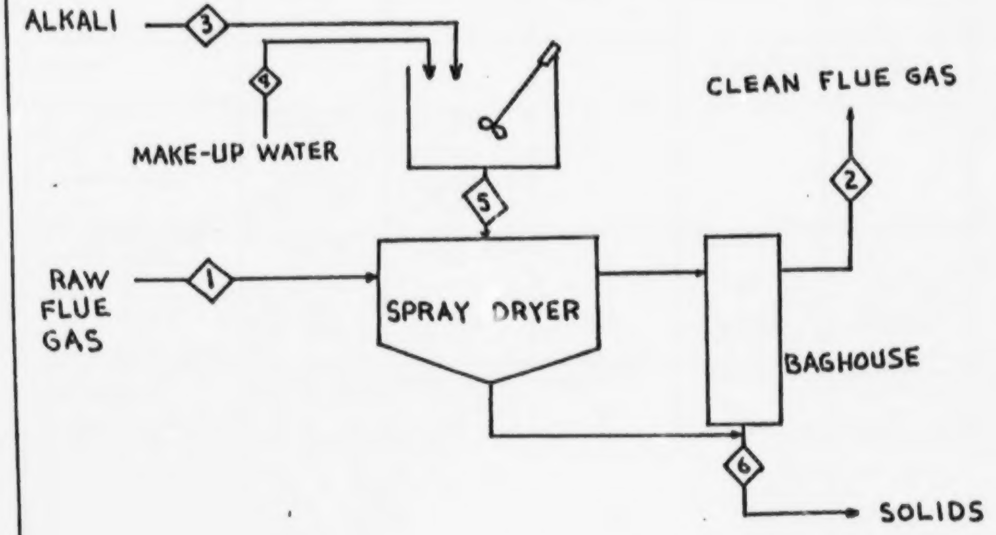
Stream Number/	1	2	3	4	5	6
Description	Inlet Gas	Exit Gas	Alkali	Water	Slurry to Dryer	Solids
TOTAL FLOW RATE						
lb/hr	185,640	190,737	311	9688	9998	1066
gpm				19.4	19.4	
Temp (°F)	350	175				
GAS FLOW RATE						
lb/hr	185,092	190,734				
lb/moles/hr	6382	6716				
ACFM	62,800	51,805				
SCFM	40,316	42,423				
LIQUID FLOW RATE						
lb/hr				9688	9688	
SOLID FLOW RATE						
lb/hr	548	3	311		311	1066



SPRAY DRYER FGD

Boiler Size: 150 x 10^6 Btu/hr
Coal Type: Western
SO_2 Removal: 75
Type Alkali: LIME

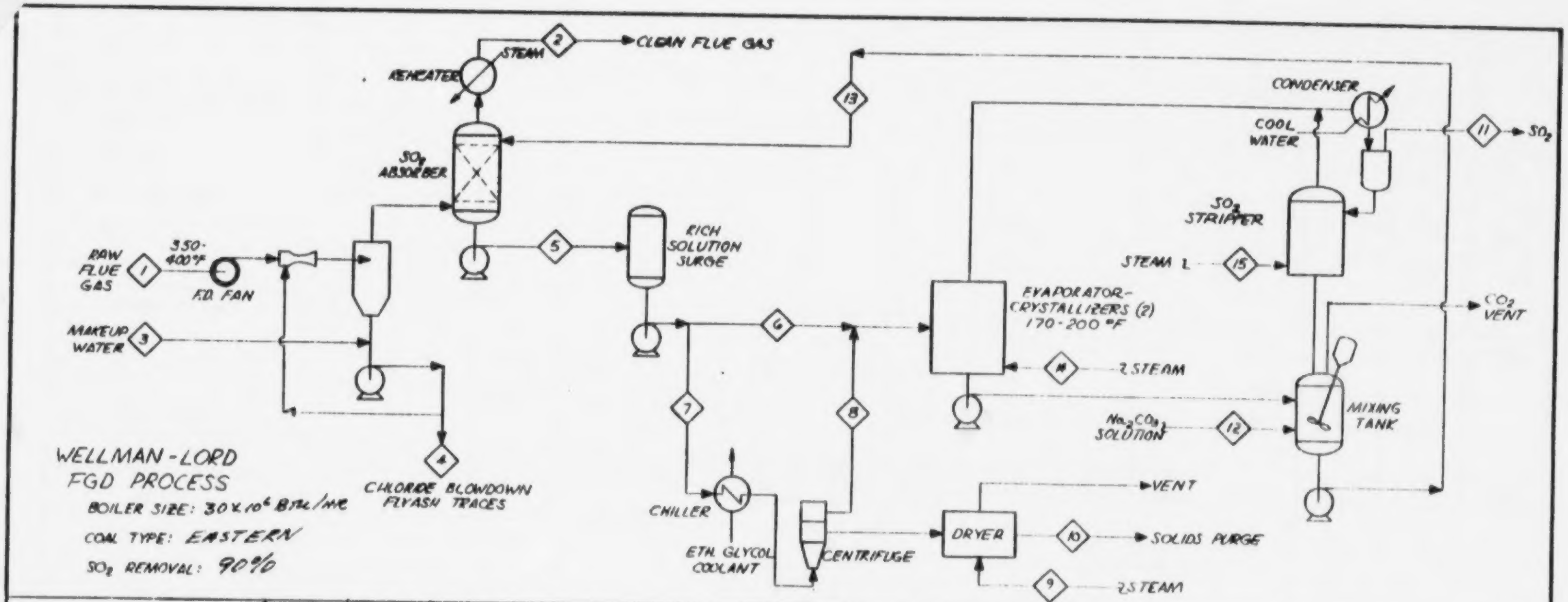
Stream Number & Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE						
lb/hr	185,640	190,709	187	9688	9875	855
gpm				19.4	19.4	
Temp (°F)	350	175				
GAS FLOW RATE						
lb/hr	185,092	190,706				
lb/moles/hr	6382	6715				
ACFM	62,800	51,797				
SCFM	40,316	42,416				
LIQUID FLOW RATE						
lb/hr				9688	9688	
SOLID FLOW RATE						
lb/hr	548	3	187		187	855



SPRAY DRYER FGD

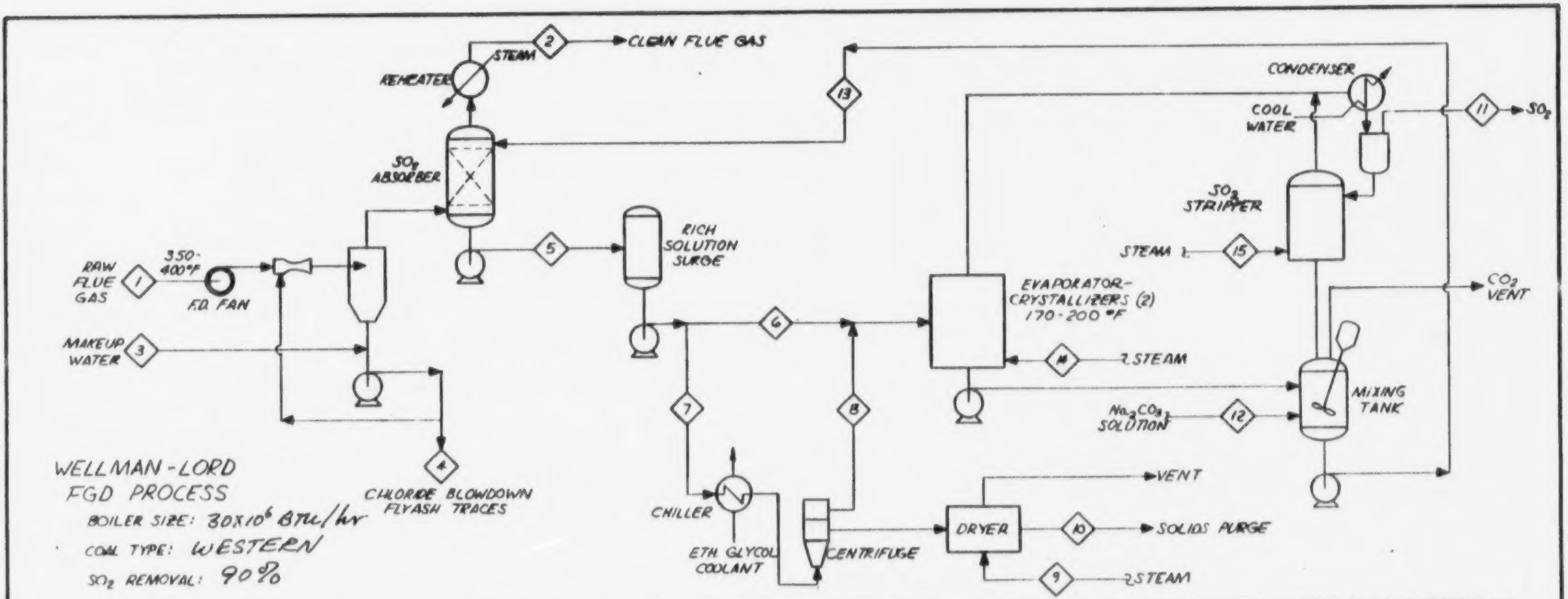
Boiler Size: 150 x10^6 Btu/hr
Coal Type: western
SO_2 Removal: 50%
Type Alkali: LIME

Stream Number & Description	1 Inlet Gas	2 Exit Gas	3 Alkali	4 Water	5 Slurry to Dryer	6 Solids
TOTAL FLOW RATE lb/hr	185,640	190,720	101	9688	9789	
gpm				19.4	19.4	
Temp (°F)	350	175				
GAS FLOW RATE lb/hr	185,092	190,717				
lb/moles/hr	6382	6715.5				
ACFM	62,800	51,800				
SCFM	40,316	42,400				
LIQUID FLOW RATE lb/hr				9688	9688	
SOLID FLOW RATE lb/hr	548	3	101		101	714



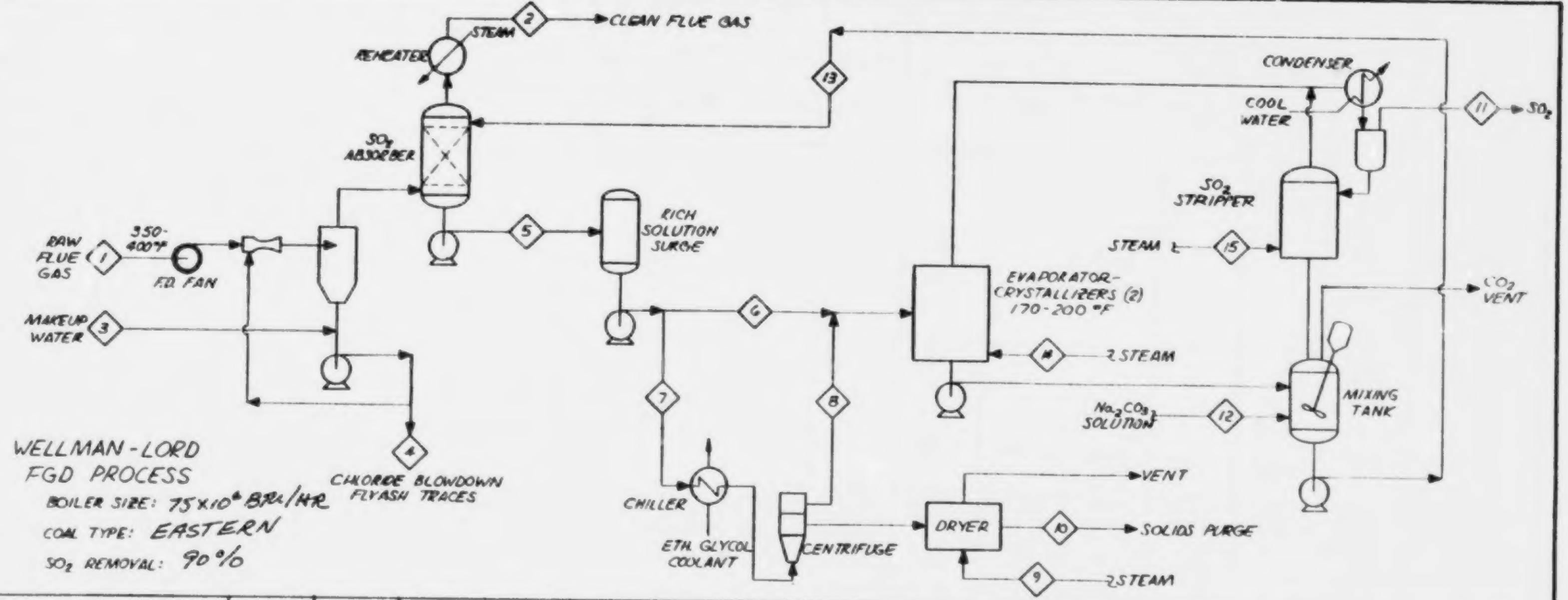
WELLMAN-LORD
FGD PROCESS
BOILER SIZE: 30 x 10⁶ BTU/HR
COAL TYPE: EASTERN
SO2 REMOVAL: 90%

STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 MAKE-UP H_2O	4 PRESCRUBBER BLOWDOWN	5 SCRUBBER BOTTOMS	6 REGENERATION FEED	7 PURGE SECTION FEED	8 PURGE SECTION RETURN	9 DRYER STEAM	10 SOLIDS PURGE	11 PRODUCT SO_2	12 Na_2CO_3	13 SCRUBBER FEED	14 REGENERATION STEAM	15 STRIPPING STEAM
TOTAL FLOW RATE															
(lb./hr.)	35831	38291	8993	1212	3161	2688	473	392	65	34.2	256	173	3115	955	173
(GPM)	--	—	17.9	2.4	9.4	3.8	0.6	0.6	—	—	—	0.3	9.8	—	—
TEMPERATURE (°F)	400	125	—	—	—	—	—	—	260	—	—	—	—	260	260
GAS FLOW RATE															
(lb./hr.)	35831	38291							65		256			955	173
(lb. MOLES/hr.)	1208	1355							3.6		8.7			53.1	9.6
(ACFM)	12900	9644													
(SCFM)	7800	8557													
LIQUID PHASE FLOW RATE															
(lb./hr.)			8993	1212	3125	2657	468	392		1.4		173	3084		
DISSOLVED SOLIDS WT. %					29.0	29.0	29.0	27.6				15.0	73.0		
SOLID PHASE FLOW RATE															
(lb./hr.)					36	31	5			32.8			3.1		



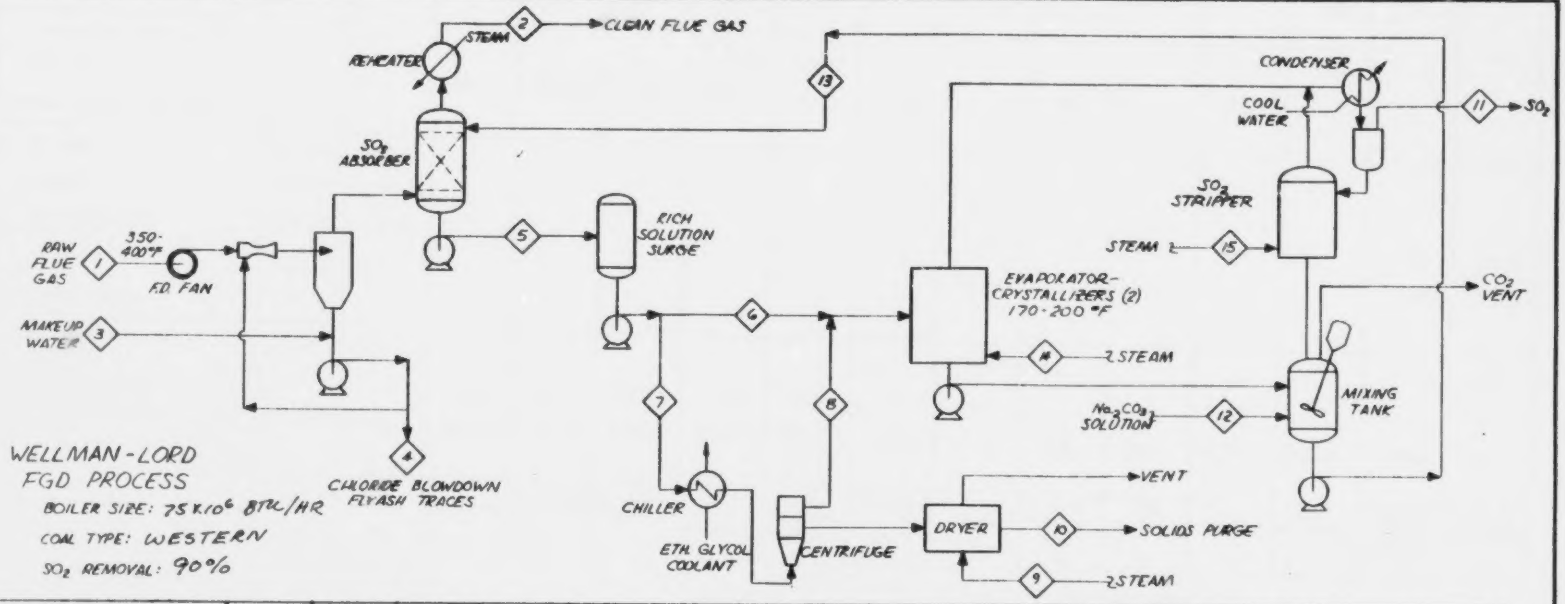
WELLMAN-LORD
FGD PROCESS
BOILER SIZE: 30x10⁶ BTU/hr
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%

STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 MAKE-UP H_2O	4 PRESCRUBBER BLOWDOWN	5 SCRUBBER BOTTOMS	6 REGENERATION FEED	7 PURGE SECTION FEED	8 PURGE SECTION RETURN	9 DRYER STEAM	10 SOLIDS PURGE	11 PRODUCT SO_2	12 Na_2CO_3	13 SCRUBBER FEED	14 REGENERATION STEAM	15 STRIPPING STEAM
TOTAL FLOW RATE															
(lb./hr.)	36977	38744	3700	1200	680	578	102	82.7	13.7	9.5	54	36.5	670	202	37
(GPM)	—	—	7.4	2.4	0.9	0.8	0.1	0.1	—	—	—	0.06	1.0	—	—
TEMPERATURE (°F)	350	125	—	—	—	—	—	—	260	—	—	—	—	260	260
GAS FLOW RATE															
(lb./hr.)	36977	38744							13.7		54			202	37
(lb. MOLES/hr.)	1253	1352							0.8		1.8			11.2	2.0
(ACFM)	12500	9621													
(SCFM)	7913	8544													
LIQUID PHASE FLOW RATE															
(lb./hr.)			3700	1200	658	559	99	82.7		0.3		36.5	651		
DISSOLVED SOLIDS WT. %					29.0	29.0	29.0	27.6				15.0	23.0		
SOLID PHASE FLOW RATE															
(lb./hr.)					22	19	3			9.2			19.0		



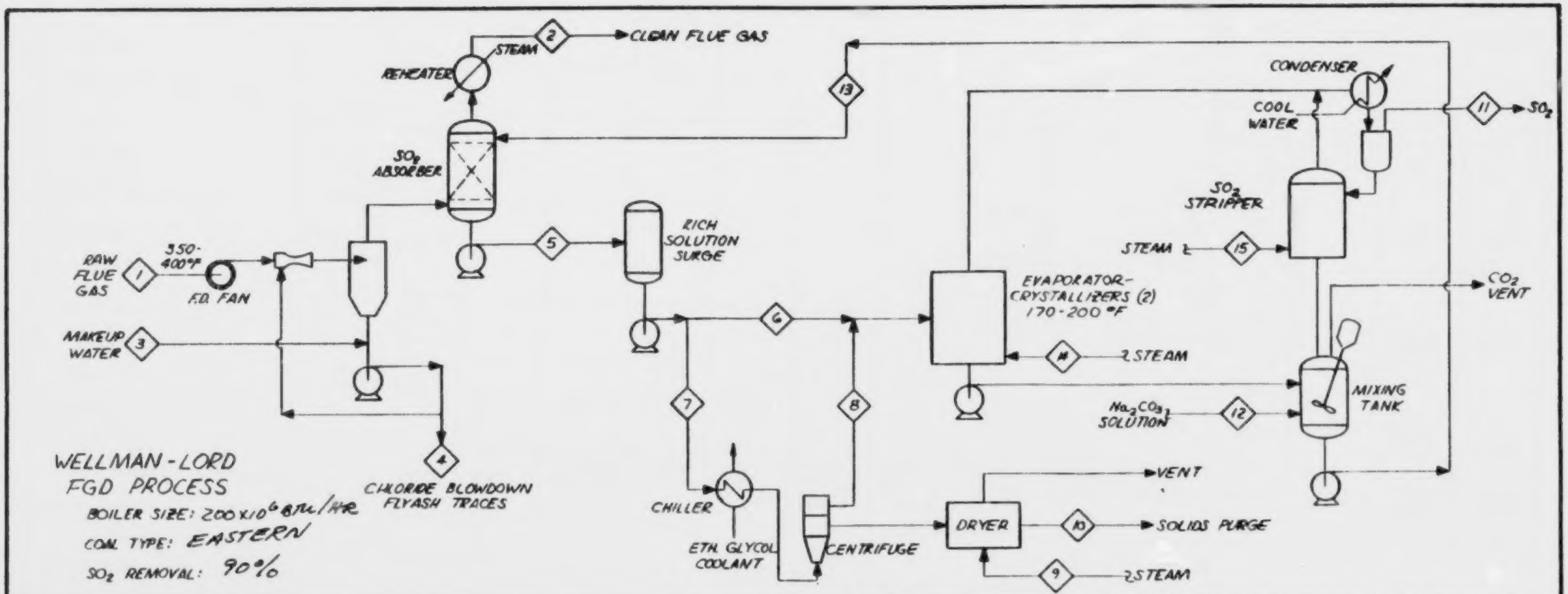
WELLMAN-LORD
FGD PROCESS
BOILER SIZE: 75×10⁶ BTU/HR
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%

STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 MAKE-UP H_2O	4 PRESCRUBBER BLOWDOWN	5 SCRUBBER BOTTOMS	6 REGENER-ATION FEED	7 PURGE SECTION FEED	8 PURGE SECTION RETURN	9 DRYER STEAM	10 SOLIDS PURGE	11 PRODUCT SO_2	12 Na_2CO_3	13 SCRUBBER FEED	14 REGENER-ATION STEAM	15 STRIPPING STEAM
TOTAL FLOW RATE															
(lb./hr.)	89576	95721	9450	3030	7902	6720	1182	980	162	83	640	432	7788	2388	432
(GPM)	—	—	19.0	6.0	11.1	94	1.7	1.4	—	–	—	0.8	11.9	—	—
TEMPERATURE (°F)	350	125	—	–	—	—	—	—	260	—	—	—		260	260
GAS FLOW RATE															
(lb./hr.)	89576	95721							162		640			2388	432
(lb. MOLES/hr.)	3020	3388							9		21.7			133	29.0
(ACFM)	31555	24110													
(SCFM)	19080	21392													
LIQUID PHASE FLOW RATE															
(lb./hr.)			9450	3030	7812	6642	1170	980		4		432	7710		
DISSOLVED SOLIDS WT. %					29.0	29.0	29.0	27.6				15.0	23.0		
SOLID PHASE FLOW RATE															
(lb./hr.)					90	78	12			79			78		



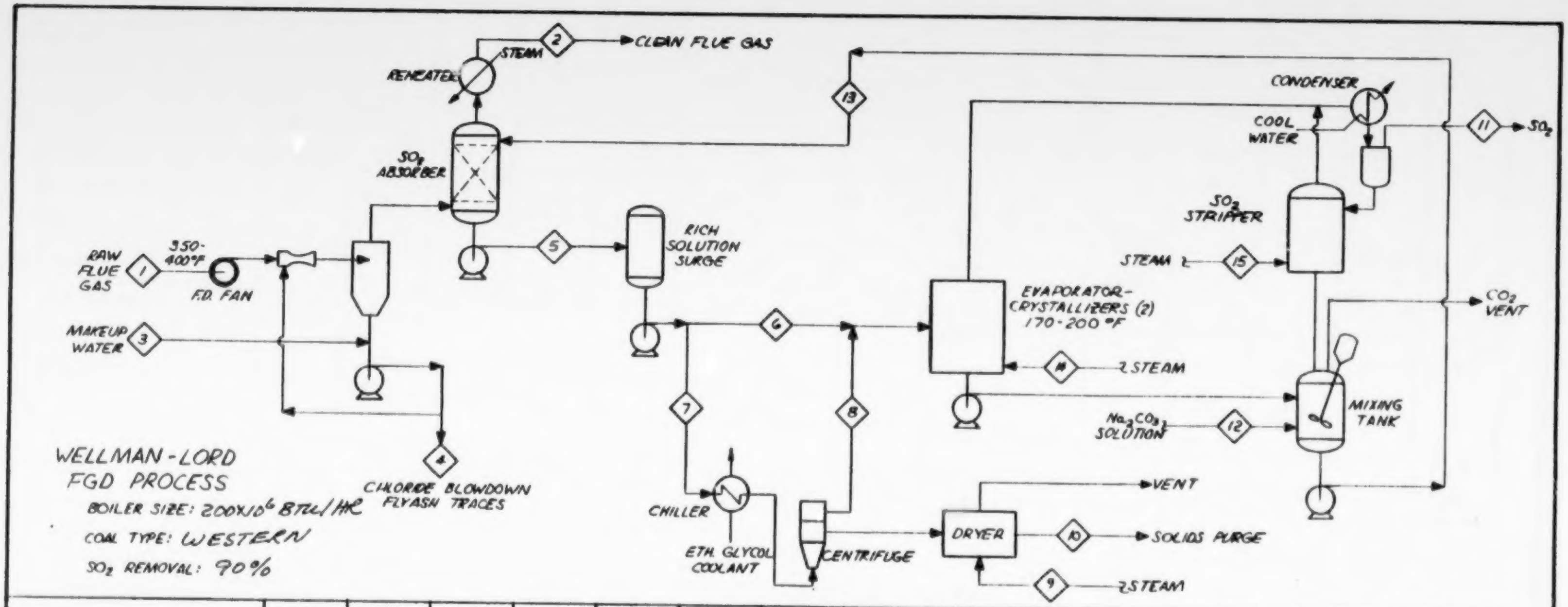
WELLMAN-LORD
FGD PROCESS
BOILER SIZE: 75 x 10^6 BTU/HR
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%

STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 MAKE-UP H_2O	4 PRESCRUBBER BLOWDOWN	5 SCRUBBER BOTTOMS	6 REGENERATION FEED	7 PURGE SECTION FEED	8 PURGE SECTION RETURN	9 DRYER STEAM	10 SOLIDS PURGE	11 PRODUCT SO_2	12 Na_2CO_3	13 SCRUBBER FEED	14 REGENERATION STEAM	15 STRIPPING STEAM
TOTAL FLOW RATE															
(lb./hr.)	92548	96880	7447	3000	1700	1445	255	207	34.3	23.7	135	91	1675	505	92.5
(GPM)	—	—	14.9	6.0	2.3	1.9	0.4	0.3	—	—	—	0.2	2.5	—	—
TEMPERATURE (°F)	350	125	—	—	—	—	—	—	260	—	—	—	—	260	260
GAS FLOW RATE															
(lb./hr.)	92548	96880							34.3		135			505	92.5
(lb. MOLES/hr.)	3132	3380							1.9		4.58			28	5.08
(ACFM)	31250	24052													
(SCFM)	19782	21345													
LIQUID PHASE FLOW RATE															
(lb./hr.)			7447	3000	1645	1398	247	207		0.8		91	1628		
DISSOLVED SOLIDS WT. %					29.0	29.0	29.0	27.6				15.0	23.0		
SOLID PHASE FLOW RATE															
(lb./hr.)					55	47	8			22.9			97.5		



WELLMAN-LORD
FGD PROCESS
BOILER SIZE: 200x10⁶ BTU/HR
COAL TYPE: EASTERN
SO_2 REMOVAL: 90%

STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 MAKE-UP H_2O	4 PRESCRUBBER BLOWDOWN	5 SCRUBBER BOTTOMS	6 REGENERATION FEED	7 PURGE SECTION FEED	8 PURGE SECTION RETURN	9 DRYER STEAM	10 SOLIDS PURGE	11 PRODUCT SO_2	12 Na_2CO_3	13 SCRUBBER FEED	14 REGENERATION STEAM	15 STRIPPING STEAM
TOTAL FLOW RATE															
(lb./hr.)	209089	222567	20492	7997	20579	18182	2447	2024	336	179	1707	893	20736	6367	1153
(GPM)	—	—	41.0	16.0	30.0	26.5	3.5	2.9	—	—	—	1.5	32.0	—	—
TEMPERATURE (°F)	400	125	—	—	—	—	—	—	260	—	—	—	—	260	260
GAS FLOW RATE															
(lb./hr.)	209089	222567							336		1707			6367	1153
(lb. MOLES/hr.)	7037	8002							18.7		57.8			354	64.1
(ACFM)	74800	56892													
(SCFM)	44450	50571													
LIQUID PHASE FLOW RATE															
(lb./hr.)			20992	7997	20372	17956	2416	2024		7		893	20560		
DISSOLVED SOLIDS WT. %					29.0	29.0	29.0	27.6				15	23.0		
SOLID PHASE FLOW RATE															
(lb./hr.)					207	176	31			172			176		



WELLMAN-LORD
FGD PROCESS
BOILER SIZE: 200×10⁶ BTU/HR
COAL TYPE: WESTERN
SO_2 REMOVAL: 90%

STREAM NUMBER & DESCRIPTION	1 INLET GAS	2 EXIT GAS	3 MAKE-UP H_2O	4 PRESCRUBBER BLOWDOWN	5 SCRUBBER BOTTOMS	6 REGENER-ATION FEED	7 PURGE SECTION FEED	8 PURGE SECTION RETURN	9 DRYER STEAM	10 SOLIDS PURGE	11 PRODUCT SO_2	12 Na_2CO_3	13 SCRUBBER FEED	14 REGENER-ATION STEAM	15 STRIPPING STEAM
TOTAL FLOW RATE															
(lb./hr.)	216643	223822	15744	7997	4217	3676	541	427	70.8	62.4	360	189	4513	1343	243
(GPM)	—	—	31.5	16.0	5.6	4.9	0.7	0.7	—	–	—	0.82	7.0	—	—
TEMPERATURE (°F)	350	125	—	—	—	—	—	—	260	–	–	–	–	260	260
GAS FLOW RATE															
(lb./hr.)	216643	223822							70.8		360			1343	243
(lb. MOLES/hr.)	7288	7713							3.92		12.2			74.7	13.5
(ACFM)	71710	52074													
(SCFM)	46036	48723													
LIQUID PHASE FLOW RATE															
(lb./hr.)			15744	7997	4217	3676	541	427		1.6		189	4513		
DISSOLVED SOLIDS WT. %					29.0	29.0	29.0	27.6		–		15	23		
SOLID PHASE FLOW RATE															
(lb./hr.)										60.8					

APPENDIX B

CAPITAL AND ANNUALIZED COST FOR FGD PROCESSES

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 30x10^6Btu/hr (8.8 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	59		
SO_2 Scrubbing	149		
Fans	80		
Wastewater Pumps			
Regeneration			
Solids Separation	109		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	20		
Total Direct Costs (TDC)		358	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	36		
Construction Fees (0.1 TDC)	36		
Start-up (0.02 TDC)	7		
Performance Test (0.01 TDC)	4		
Total Indirect Cost (TIC)		166	
Contingencies (0.2 (TDC+TIC))		105	
TOTAL TURNKEY COSTS (TTC)			629
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		62	
TOTAL CAPITAL INVESTMENT (TCI)			691

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200x10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400x10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 30x10^{6}Btu/hr (8.8 MW$_t$)
Coal Feedstock: Eastern 3.5% S
SO$_2$ Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			14		
Maintenance Materials (.04 TDC)			14		
Electricity (25.8 mills/kWh)	54.7	kW	7		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/m^3)	1.4	m^3/hr	0.2		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	323	kg/hr	75		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	127	kg/hr	10		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				246	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			40		
Total Overhead Costs				78	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				117	
TOTAL ANNUALIZED COSTS					441
Annual Unit Costs	$/10^6 Btu		$/kg SO$_2$		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: $30x10^6$ Btu/hr (8.8 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 85%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	57		
SO_2 Scrubbing	142		
Fans	20		
Wastewater Pumps			
Regeneration			
Solids Separation	107		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	20		
Total Direct Costs (TDC)		346	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	35		
Construction Fees (0.1 TDC)	35		
Start-up (0.02 TDC)	7		
Performance Test (0.01 TDC)	3		
Total Indirect Cost (TIC)		163	
Contingencies (0.2 (TDC+TIC))		102	
TOTAL TURNKEY COSTS (TTC)			611
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		60	
TOTAL CAPITAL INVESTMENT (TCI)			671

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 30×10^6 Btu/hr (8.8 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 85%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			14		
Maintenance Materials (.04 TDC)			14		
Electricity (25.8 mills/kWh)	45	kW	6		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	1.36	m^3/hr	0.3		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	305	kg/hr	70		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	120	kg/hr	9		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				239	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			40		
Total Overhead Costs				78	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				114	
TOTAL ANNUALIZED COSTS					431
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 30×10^6 Btu/hr (8.8 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	55		
SO_2 Scrubbing	131		
Fans	20		
Wastewater Pumps			
Regeneration			
Solids Separation	106		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	19		
Total Direct Costs (TDC)		332	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	33		
Construction Fees (0.1 TDC)	33		
Start-up (0.02 TDC)	7		
Performance Test (0.01 TDC)	3		
Total Indirect Cost (TIC)		159	
Contingencies (0.2 (TDC+TIC))		98	
TOTAL TURNKEY COSTS (TTC)			589
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		57	
TOTAL CAPITAL INVESTMENT (TCI)			646

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 30×10^6 Btu/hr (8.8 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			13		
Maintenance Materials (.04 TDC)			13		
Electricity (25.8 mills/kWh)	35.9	kW	5		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	1.34	m^3/hr	0.3		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	270	kg/hr	63		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	106	kg/hr	8		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				228	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			40		
Total Overhead Costs				78	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				110	
TOTAL ANNUALIZED COSTS					416
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 75×10^6 Btu/hr (22 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	99		
SO_2 Scrubbing	244		
Fans	40		
Wastewater Pumps			
Regeneration			
Solids Separation	121		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	30		
Total Direct Costs (TDC)		534	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	53		
Construction Fees (0.1 TDC)	53		
Start-up (0.02 TDC)	11		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		205	
Contingencies (0.2 (TDC+TIC))		148	
TOTAL TURNKEY COSTS (TTC)			887
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		100	
TOTAL CAPITAL INVESTMENT (TCI)			987

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: $75\text{x}10^6$ Btu/hr (22 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			21		
Maintenance Materials (.04 TDC)			19		
Electricity (25.8 mills/kWh)	137	kW			
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	3.5	m^3/hr	0.7		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	816	kg/hr	188		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	322	kg/hr	24		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				400	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			44		
Total Overhead Costs				82	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				168	
TOTAL ANNUALIZED COSTS					650
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 75x10^6 Btu/hr (22 MW$_t$)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	89		
SO_2 Scrubbing	216		
Fans	40		
Wastewater Pumps			
Regeneration			
Solids Separation	118		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	28		
Total Direct Costs (TDC)		490	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	49		
Construction Fees (0.1 TDC)	49		
Start-up (0.02 TDC)	10		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		196	
Contingencies (0.2 (TDC+TIC))		137	
TOTAL TURNKEY COSTS (TTC)			823
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		89	
TOTAL CAPITAL INVESTMENT (TCI)			913

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200x10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400x10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 75×10^6 Btu/hr (22 MW_t)
Coal Feedstock: Fastern 3.5% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			20		
Maintenance Materials (.04 TDC)			20		
Electricity (25.8 mills/kWh)	86	kW	12		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	3.4	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	685	kg/hr	158		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	270	kg/hr	20		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				357	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			43		
Total Overhead Costs				81	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				155	
TOTAL ANNUALIZED COSTS					593
Annual Unit Costs	$/$10^6$ Btu		\$/kg SO_2		

---

a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	147		
SO_2 Scrubbing	368		
Fans	69		
Wastewater Pumps			
Regeneration			
Solids Separation	133		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	43		
Total Direct Costs (TDC)		760	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	76		
Construction Fees (0.1 TDC)	76		
Start-up (0.02 TDC)	15		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		258	
Contingencies (0.2 (TDC+TIC))		204	
TOTAL TURNKEY COSTS (TTC)			1221
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		163	
TOTAL CAPITAL INVESTMENT (TCI)			1385

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			30		
Maintenance Materials (.04 TDC)			30		
Electricity (25.8 mills/kWh)	265	kW	36		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	6.9	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	1647	kg/hr	381		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	647	kg/hr	49		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				653	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			48		
Total Overhead Costs				86	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				235	
TOTAL ANNUALIZED COSTS					974
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone

Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)

Coal Feedstock: Eastern 3.5% S

SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	134		
SO_2 Scrubbing	325		
Fans	69		
Wastewater Pumps			
Regeneration			
Solids Separation	129		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	39		
Total Direct Costs (TDC)		696	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	70		
Construction Fees (0.1 TDC)	70		
Start-up (0.02 TDC)	14		
Performance Test (0.01 TDC)	7		
Total Indirect Cost (TIC)		244	
Contingencies (0.2 (TDC+TIC))		188	
TOTAL TURNKEY COSTS (TTC)			1128
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		141	
TOTAL CAPITAL INVESTMENT (TCI)			1270

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			28		
Maintenance Materials (.04 TDC)			28		
Electricity (25.8 mills/kWh)	173	kW	23		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	6.9	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	1372	kg/hr	318		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	540	kg/hr	40		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				564	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			47		
Total Overhead Costs				85	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				216	
TOTAL ANNUALIZED COSTS					865
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200×10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	171		
SO_2 Scrubbing	401		
Fans	76		
Wastewater Pumps			
Regeneration			
Solids Separation	137		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	47		
Total Direct Costs (TDC)		832	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	83		
Construction Fees (0.1 TDC)	83		
Start-up (0.02 TDC)	17		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		274	
Contingencies (0.2 (TDC+TIC))		221	
TOTAL TURNKEY COSTS (TTC)			1327
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		202	
TOTAL CAPITAL INVESTMENT (TCI)			1530

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200x10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			33		
Maintenance Materials (.04 TDC)			33		
Electricity (25.8 mills/kWh)	311	kW	42		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	8.1	m^3/hr	2		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	2187	kg/hr	506		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	862	kg/hr	65		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				807	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			50		
Total Overhead Costs				88	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				260	
TOTAL ANNUALIZED COSTS					1155
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 20x10^{6}Btu/hr (58.6 MW)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 85%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	166		
SO_2 Scrubbing	375		
Fans	76		
Wastewater Pumps			
Regeneration			
Solids Separation	135		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	45		
Total Direct Costs (TDC)		797	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	80		
Construction Fees (0.1 TDC)	80		
Start-up (0.02 TDC)	16		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		267	
Contingencies (0.2 (TDC+TIC))		213	
TOTAL TURNKEY COSTS (TTC)			1277
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		192	
TOTAL CAPITAL INVESTMENT (TCI)			1470

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200x10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400x10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200x10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 85%
Operating Factor: 60 %

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			32		
Maintenance Materials (.04 TDC)			32		
Electricity (25.8 mills/kWh)	25.4	kW	34		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	8.2	m^3/hr	2		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	2076	kg/hr	479		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	818	kg/hr	61		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				766	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			49		
Total Overhead Costs				87	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				250	
TOTAL ANNUALIZED COSTS					1102
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200×10^6 Btu/hr (58.6 MW)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	155		
SO_2 Scrubbing	354		
Fans	76		
Wastewater Pumps			
Regeneration			
Solids Separation	131		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	42		
Total Direct Costs (TDC)		757	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	76		
Construction Fees (0.1 TDC)	76		
Start-up (0.02 TDC)	15		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		258	
Contingencies (0.2 (TDC+TIC))		203	
TOTAL TURNKEY COSTS (TTC)			1218
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		173	
TOTAL CAPITAL INVESTMENT (TCI)			1392

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200×10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Eastern 3.5% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			30		
Maintenance Materials (.04 TDC)			30		
Electricity (25.8 mills/kWh)	205	kW	28		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	7.9	m^3/hr	2		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	1827	kg/hr	422		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	720	kg/hr	54		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				692	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			48		
Total Overhead Costs				86	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				236	
TOTAL ANNUALIZED COSTS					1014
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 30×10^6 Btu/hr (8.8 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	33		
SO_2 Scrubbing	142		
Fans	20		
Wastewater Pumps			
Regeneration			
Solids Separation	94		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	17		
Total Direct Costs (TDC)		305	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	30		
Construction Fees (0.1 TDC)	30		
Start-up (0.02 TDC)	6		
Performance Test (0.01 TDC)	3		
Total Indirect Cost (TIC)		152	
Contingencies (0.2 (TDC+TIC))		91	
TOTAL TURNKEY COSTS (TTC)			548
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		46	
TOTAL CAPITAL INVESTMENT (TCI)			594

[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: $30\text{x}10^6$ Btu/hr (8.8 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		12		
Maintenance Materials (.04 TDC)		12		
Electricity (25.8 mills/kWh)	42 kW	6		
Steam ($1.84/GJ)	GJ/hr			
Proc. Water ($.04/m^3)	1.2 m^3/hr	0.2		
Methane ($2.05/GJ)	GJ/hr			
Wastewater Treating[a]	m^3/hr			
Solids Disposal ($.044/kg)	75 kg/hr	18		
Chemicals				
Lime ($.0385/kg)	kg/hr			
Limestone ($.0143/kg)	27 kg/hr	2		
Na_2CO_3 ($.0991/kg)	kg/hr			
Total Direct Operating Cost			186	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		38		
Total Overhead Costs			76	
By-Product Credits			(0)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			100	
TOTAL ANNUALIZED COSTS				362
Annual Unit Costs	$/10^6 Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FCD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone

Boiler Capacity: 30x10 Btu/hr (8.8 MW$_t$)

Coal Feedstock: Western 0.6% S

SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	31		
SO_2 Scrubbing	128		
Fans	20		
Wastewater Pumps			
Regeneration			
Solids Separation	94		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	16		
Total Direct Costs (TDC)		289	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	29		
Construction Fees (0.1 TDC)	29		
Start-up (0.02 TDC)	6		
Performance Test (0.01 TDC)	3		
Total Indirect Cost (TIC)		150	
Contingencies (0.2 (TDC+TIC))		88	
TOTAL TURNKEY COSTS (TTC)			527
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		42	
TOTAL CAPITAL INVESTMENT (TCI)			569

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200x10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400x10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 30×10^6 Btu/hr
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			12		
Maintenance Materials (.04 TDC)			12		
Electricity (25.8 mills/kWh)	31	kW	4		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/m^3)	1.0	m^3/hr	0.2		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	63	kg/hr	14		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	23	kg/hr	2		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				170	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			39		
Total Overhead Costs				87	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				97	
TOTAL ANNUALIZED COSTS					
Annual Unit Costs	$/10^6 Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 30×10^6 Btu/hr (8.8 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 85%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	32		
SO_2 Scrubbing	136		
Fans	20		
Wastewater Pumps			
Regeneration			
Solids Separation	94		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	17		
Total Direct Costs (TDC)		298	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	30		
Construction Fees (0.1 TDC)	30		
Start-up (0.02 TDC)	6		
Performance Test (0.01 TDC)	3		
Total Indirect Cost (TIC)		152	
Contingencies (0.2 (TDC+TIC))		90	
TOTAL TURNKEY COSTS (TTC)			540
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		43	
TOTAL CAPITAL INVESTMENT (TCI)			583

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 30×10^6 Btu/hr (8.8 MW_t)
Coal Feedstock: ~~Western 0.6% S~~
SO_2 Control Level: 85%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		12		
Maintenance Materials (.04 TDC)		12		
Electricity (25.8 mills/kWh)	35 kW	5		
Steam ($1.84/GJ)	GJ/hr			
Proc. Water ($.04/$m^3$)	1.1 m^3/hr	0.2		
Methane ($2.05/GJ)	GJ/hr			
Wastewater Treating[a]	m^3/hr			
Solids Disposal ($.044/kg)	70 kg/hr	16		
Chemicals				
Lime ($.0385/kg)	kg/hr			
Limestone ($.0143/kg)	25 kg/hr	2		
Na_2CO_3 ($.0991/kg)	kg/hr			
Total Direct Operating Cost			173	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		39		
Total Overhead Costs			87	
By-Product Credits			(0)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			99	
TOTAL ANNUALIZED COSTS				359
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: $75\text{x}10^6$ Btu/hr (22MW$_t$)
Coal Feedstock: Western 0.6% S
SO$_2$ Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	48		
SO$_2$ Scrubbing	233		
Fans	40		
Wastewater Pumps			
Regeneration			
Solids Separation	102		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	25		
Total Direct Costs (TDC)		448	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	45		
Construction Fees (0.1 TDC)	45		
Start-up (0.02 TDC)	9		
Performance Test (0.01 TDC)	4		
Total Indirect Cost (TIC)		186	
Contingencies (0.2 (TDC+TIC))		127	
TOTAL TURNKEY COSTS (TTC)			761
Land[b] (0.00084 TTC)		0.6	
Working Capital (0.25 Direct Operating Costs)[c]		60	
TOTAL CAPITAL INVESTMENT (TCI)			821

[a] Engineering Costs = 0.1 TDC for the 58.6 MW ($200\text{x}10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO$_2$ removal. This cost remains constant for the smaller cases. For the 118 MW ($400\text{x}10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO$_2$ removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 75×10^6 Btu/hr (22 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		18		
Maintenance Materials (.04 TDC)		18		
Electricity (25.8 mills/kWh)	106 kW	14		
Steam ($1.84/GJ)	GJ/hr			
Proc. Water ($.04/$m^3$)	2.7 m^3/hr	1		
Methane ($2.05/GJ)	GJ/hr			
Wastewater Treating[a]	m^3/hr			
Solids Disposal ($.044/kg)	188 kg/hr	57		
Chemicals				
Lime ($.0385/kg)	kg/hr			
Limestone ($.0143/kg)	68 kg/hr	5		
Na_2CO_3 ($.0991/kg)	kg/hr			
Total Direct Operating Cost			239	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		42		
Total Overhead Costs			80	
By-Product Credits			(0)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			139	
TOTAL ANNUALIZED COSTS				458
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 75×10^6 Btu/hr (22 MW)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	44		
SO_2 Scrubbing	211		
Fans	40		
Wastewater Pumps			
Regeneration			
Solids Separation	101		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	24		
Total Direct Costs (TDC)		420	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	42		
Construction Fees (0.1 TDC)	42		
Start-up (0.02 TDC)	8		
Performance Test (0.01 TDC)	4		
Total Indirect Cost (TIC)		179	
Contingencies (0.2 (TDC+TIC))		120	
TOTAL TURNKEY COSTS (TTC)			719
Land[b] (0.00084 TTC)		0.6	
Working Capital (0.25 Direct Operating Costs)[c]		52	
TOTAL CAPITAL INVESTMENT (TCI)			771

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: $75\text{x}10^6$ Btu/hr (22 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			17		
Maintenance Materials (.04 TDC)			17		
Electricity (25.8 mills/kWh)	77	kW	10		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	2.7	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	157	kg/hr	36		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	57	kg/hr	4		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				210	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			42		
Total Overhead Costs				80	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				131	
TOTAL ANNUALIZED COSTS					421
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	67		
SO_2 Scrubbing	351		
Fans	66		
Wastewater Pumps			
Regeneration			
Solids Separation	112		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	36		
Total Direct Costs (TDC)		632	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	63		
Construction Fees (0.1 TDC)	63		
Start-up (0.02 TDC)	13		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		228	
Contingencies (0.2 (TDC+TIC))		172	
TOTAL TURNKEY COSTS (TTC)			1032
Land[b] (0.00084 TTC)		0.8	
Working Capital (0.25 Direct Operating Costs)[c]		76	
TOTAL CAPITAL INVESTMENT (TCI)			1108

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		25		
Maintenance Materials (.04 TDC)		25		
Electricity (25.8 mills/kWh)	212 kW	29		
Steam ($1.84/GJ)	GJ/hr			
Proc. Water ($.04/$m^3$)	5.4 m^3/hr	1		
Methane ($2.05/GJ)	GJ/hr			
Wastewater Treating[a]	m^3/hr			
Solids Disposal ($.044/kg)	377 kg/hr	88		
Chemicals				
Lime ($.0385/kg)	kg/hr			
Limestone ($.0143/kg)	136 kg/hr	10		
Na_2CO_3 ($.0991/kg)	kg/hr			
Total Direct Operating Cost			304	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		46		
Total Overhead Costs			84	
By-Product Credits			(0)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			188	
TOTAL ANNUALIZED COSTS				576
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

---

a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	62		
SO_2 Scrubbing	319		
Fans	66		
Wastewater Pumps			
Regeneration			
Solids Separation	111		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	33		
Total Direct Costs (TDC)		591	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	59		
Construction Fees (0.1 TDC)	59		
Start-up (0.02 TDC)	12		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		219	
Contingencies (0.2 (TDC+TIC))		162	
TOTAL TURNKEY COSTS (TTC)			972
Land[b] (0.00084 TTC)		0.8	
Working Capital (0.25 Direct Operating Costs)[c]		69	
TOTAL CAPITAL INVESTMENT (TCI)			1041

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200x10 Btu/hr (58.6 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	75		
SO_2 Scrubbing	384		
Fans	73		
Wastewater Pumps			
Regeneration			
Solids Separation	116		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	39		
Total Direct Costs (TDC)		686	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	69		
Construction Fees (0.1 TDC)	69		
Start-up (0.02 TDC)	14		
Performance Test (0.01 TDC)	7		
Total Indirect Cost (TIC)		242	
Contingencies (0.2 (TDC+TIC))		186	
TOTAL TURNKEY COSTS (TTC)			1114
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		86	
TOTAL CAPITAL INVESTMENT (TCI)			1201

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a. Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200x10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			27		
Maintenance Materials (.04 TDC)			27		
Electricity (25.8 mills/kWh)	247	kW	33		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	5	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	503	kg/hr	117		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	182	kg/hr	14		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				345	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			47		
Total Overhead Costs				85	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				204	
TOTAL ANNUALIZED COSTS					634
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		24		
Maintenance Materials (.04 TDC)		24		
Electricity (25.8 mills/kWh)	153 kW	21		
Steam ($1.84/GJ)	GJ/hr			
Proc. Water ($.04/$m^3$)	5.4 m^3/hr	1		
Methane ($2.05/GJ)	GJ/hr			
Wastewater Treating[a]	m^3/hr			
Solids Disposal ($.044/kg)	314 kg/hr	72		
Chemicals				
Lime ($.0385/kg)	kg/hr			
Limestone ($.0143/kg)	114 kg/hr	9		
Na_2CO_3 ($.0991/kg)	kg/hr			
Total Direct Operating Cost			277	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		45		
Total Overhead Costs			83	
By-Product Credits			(0)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			177	
TOTAL ANNUALIZED COSTS				537
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200×10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 85%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	73		
SO_2 Scrubbing	367		
Fans	73		
Wastewater Pumps			
Regeneration			
Solids Separation	116		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	38		
Total Direct Costs (TDC)		666	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	67		
Construction Fees (0.1 TDC)	67		
Start-up (0.02 TDC)	13		
Performance Test (0.01 TDC)	7		
Total Indirect Cost (TIC)		237	
Contingencies (0.2 (TDC+TIC))		181	
TOTAL TURNKEY COSTS (TTC)			1084
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		83	
TOTAL CAPITAL INVESTMENT (TCI)			1168

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200×10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 85%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			27		
Maintenance Materials (.04 TDC)			27		
Electricity (25.8 mills/kWh)	208	kW	28		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	5	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	475	kg/hr	109		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	172	kg/hr	13		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				331	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			47		
Total Overhead Costs				85	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				198	
TOTAL ANNUALIZED COSTS					614
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200×10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	69		
SO_2 Scrubbing	348		
Fans	73		
Wastewater Pumps			
Regeneration			
Solids Separation	116		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	36		
Total Direct Costs (TDC)		642	
Indirect Capital Costs			
Engineering[a]	83		
Construction and Field Expenses (0.1 TDC)	64		
Construction Fees (0.1 TDC)	64		
Start-up (0.02 TDC)	13		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		230	
Contingencies (0.2 (TDC+TIC))		174	
TOTAL TURNKEY COSTS (TTC)			1046
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		78	
TOTAL CAPITAL INVESTMENT (TCI)			1125

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Limestone
Boiler Capacity: 200×10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Western 0.6 % S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			26		
Maintenance Materials (.04 TDC)			26		
Electricity (25.8 mills/kWh)	175	kW	24		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	5	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	418	kg/hr	97		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)	151	kg/hr	11		
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				311	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			46		
Total Overhead Costs				84	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				191	
TOTAL ANNUALIZED COSTS					586
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway

Boiler Capacity: 8.8 MW_t (30 MBtu/hr

Coal Feedstock: 3.5% S Eastern

SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	63		
SO_2 Scrubbing	109		
Fans	20		
Wastewater Pumps	18		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	13		
Total Direct Costs (TDC)		223	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	22		
Construction Fees (0.1 TDC)	22		
Start-up (0.02 TDC)	4		
Performance Test (0.01 TDC)	2		
Total Indirect Cost (TIC)		112	
Contingencies (0.2 (TDC+TIC))		67	
TOTAL TURNKEY COSTS (TTC)			402
Land[b] (0.00084 TTC)		0.3	
Working Capital (0.25 Direct Operating Costs)[c]		55	
TOTAL CAPITAL INVESTMENT (TCI)			457

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		9		
Maintenance Materials (.04 TDC)		9		
Electricity (25.8 mills/kWh)	24.3 kW	3		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	4.25 m^3/hr	1		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	3.04 m^3/hr	5		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	131 kg/hr	68		
Total Direct Operating Cost			221	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		37		
Total Overhead Costs			75	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			78	
TOTAL ANNUALIZED COSTS				374
Annual Unit Costs	$/10^6 Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 85%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	63		
SO_2 Scrubbing	109		
Fans	20		
Wastewater Pumps	18		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	13		
Total Direct Costs (TDC)		223	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	22		
Construction Fees (0.1 TDC)	22		
Start-up (0.02 TDC)	4		
Performance Test (0.01 TDC)	2		
Total Indirect Cost (TIC)		112	
Contingencies (0.2 (TDC+TIC))		67	
TOTAL TURNKEY COSTS (TTC)			402
Land[b] (0.00084 TTC)		0.3	
Working Capital (0.25 Direct Operating Costs)[c]		55	
TOTAL CAPITAL INVESTMENT (TCI)			456

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 85%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		9		
Maintenance Materials (.04 TDC)		9		
Electricity (25.8 mills/kWh)	23.7 kW	3		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	4.04 m^3/hr	1		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	2.84 m^3/hr	5		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	125 kg/hr	65		
Total Direct Operating Cost			218	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		37		
Total Overhead Costs			75	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			78	
TOTAL ANNUALIZED COSTS				371
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	60		
SO_2 Scrubbing	109		
Fans	20		
Wastewater Pumps	17		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	12		
Total Direct Costs (TDC)		218	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	22		
Construction Fees (0.1 TDC)	22		
Start-up (0.02 TDC)	4		
Performance Test (0.01 TDC)	2		
Total Indirect Cost (TIC)		112	
Contingencies (0.2 (TDC+TIC))		66	
TOTAL TURNKEY COSTS (TTC)			396
Land[b] (0.00084 TTC)		0.3	
Working Capital (0.25 Direct Operating Costs)[c]		53	
TOTAL CAPITAL INVESTMENT (TCI)			449

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 75%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		9		
Maintenance Materials (.04 TDC)		9		
Electricity (25.8 mills/kWh)	22.9 kW	3		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/m³)	3.75 m³/hr	1		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	2.54 m³/hr	4		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	116 kg/hr	60		
Total Direct Operating Cost			212	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		37		
Total Overhead Costs			75	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			76	
TOTAL ANNUALIZED COSTS				363
Annual Unit Costs	$/10⁶ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	104		
SO_2 Scrubbing	182		
Fans	40		
Wastewater Pumps	21		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	21		
Total Direct Costs (TDC)		368	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	37		
Construction Fees (0.1 TDC)	37		
Start-up (0.02 TDC)	7		
Performance Test (0.01 TDC)	4		
Total Indirect Cost (TIC)		147	
Contingencies (0.2 (TDC+TIC))		103	
TOTAL TURNKEY COSTS (TTC)			618
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		87	
TOTAL CAPITAL INVESTMENT (TCI)			705

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		15		
Maintenance Materials (.04 TDC)		15		
Electricity (25.8 mills/kWh)	61.7 kW	8		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	10.9 m^3/hr	2		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	7.85 m^3/hr	7		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	337 kg/hr	175		
Total Direct Operating Cost			348	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		41		
Total Overhead Costs			79	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			120	
TOTAL ANNUALIZED COSTS				547
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	96		
SO_2 Scrubbing	182		
Fans	40		
Wastewater Pumps	21		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	20		
Total Direct Costs (TDC)		359	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	36		
Construction Fees (0.1 TDC)	36		
Start-up (0.02 TDC)	7		
Performance Test (0.01 TDC)	4		
Total Indirect Cost (TIC)		145	
Contingencies (0.2 (TDC+TIC))		101	
TOTAL TURNKEY COSTS (TTC)			605
Land[b] (0.00084 TTC)		0.4	
Working Capital (0.25 Direct Operating Costs)[c]		80	
TOTAL CAPITAL INVESTMENT (TCI)			685

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 75%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		14		
Maintenance Materials (.04 TDC)		14		
Electricity (25.8 mills/kWh)	57.5 kW	8		
Steam ($1.84/GJ)	GJ/hr	–		
Proc. Water ($.04/$m^3$)	9.42 m^3/hr	2		
Methane ($2.05/GJ)	GJ/hr	–		
Wastewater Treating[a]	6.38 m^3/hr	6		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	–		
Limestone ($.0143/kg)	kg/hr	–		
Na_2CO_3 ($.0991/kg)	291 kg/hr	151		
Total Direct Operating Cost			321	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		40		
Total Overhead Costs			78	
By-Product Credits			(–)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			116	
TOTAL ANNUALIZED COSTS				515
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 44 MW_t (150 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	154		
SO_2 Scrubbing	275		
Fans	68		
Wastewater Pumps	25		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	31		
Total Direct Costs (TDC)		553	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	55		
Construction Fees (0.1 TDC)	55		
Start-up (0.02 TDC)	11		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		189	
Contingencies (0.2 (TDC+TIC))		148	
TOTAL TURNKEY COSTS (TTC)			890
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		138	
TOTAL CAPITAL INVESTMENT (TCI)			1028

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a. Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 44 MW_t (150 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		22		
Maintenance Materials (.04 TDC)		22		
Electricity (25.8 mills/kWh)	124 kW	17		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	21.8 m^3/hr	5		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	15.8 m^3/hr	10		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	675 kg/hr	352		
Total Direct Operating Cost			554	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		44		
Total Overhead Costs			82	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			175	
TOTAL ANNUALIZED COSTS				811
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 44 MW_t (150 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	142		
SO_2 Scrubbing	275		
Fans	68		
Wastewater Pumps	24		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	30		
Total Direct Costs (TDC)		539	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	54		
Construction Fees (0.1 TDC)	54		
Start-up (0.02 TDC)	11		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		186	
Contingencies (0.2 (TDC+TIC))		143	
TOTAL TURNKEY COSTS (TTC)			868
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		125	
TOTAL CAPITAL INVESTMENT (TCI)			993

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 44 MW_t (150 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			22		
Maintenance Materials (.04 TDC)			22		
Electricity (25.8 mills/kWh)	115	kW	16		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	18.8	m^3/hr	4		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	12.7	m^3/hr	9		
Solids Disposal ($.044/kg)		kg/hr			
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	579	kg/hr	302		
Total Direct Operating Cost				501	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			44		
Total Overhead Costs				82	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				169	
TOTAL ANNUALIZED COSTS					752
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	178		
SO_2 Scrubbing	301		
Fans	76		
Wastewater Pumps	26		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	35		
Total Direct Costs (TDC)		616	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	62		
Construction Fees (0.1 TDC)	62		
Start-up (0.02 TDC)	12		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		204	
Contingencies (0.2 (TDC+TIC))		164	
TOTAL TURNKEY COSTS (TTC)			984
Land[b] (0.00084 TTC)		0.8	
Working Capital (0.25 Direct Operating Costs)[c]		167	
TOTAL CAPITAL INVESTMENT (TCI)			1151

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			27		
Maintenance Materials (.04 TDC)			27		
Electricity (25.8 mills/kWh)	151	kW	20		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	27.9	m^3/hr	6		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	20.9	m^3/hr	11		
Solids Disposal ($.044/kg)		kg/hr			
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	864	kg/hr	450		
Total Direct Operating Cost				667	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			47		
Total Overhead Costs				85	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				196	
TOTAL ANNUALIZED COSTS					947
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 85%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	171		
SO_2 Scrubbing	301		
Fans	76		
Wastewater Pumps	26		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	34		
Total Direct Costs (TDC)		608	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	61		
Construction Fees (0.1 TDC)	61		
Start-up (0.02 TDC)	12		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		202	
Contingencies (0.2 (TDC+TIC))		162	
TOTAL TURNKEY COSTS (TTC)			972
Land[b] (0.00084 TTC)		0.8	
Working Capital (0.25 Direct Operating Costs)[c]		157	
TOTAL CAPITAL INVESTMENT (TCI)			1129

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.55 S Eastern
SO_2 Control Level: 85%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			24		
Maintenance Materials (.04 TDC)			24		
Electricity (25.8 mills/kWh)	145	kW	20		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	26.0	m^3/hr	5		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	18.9	m^3/hr	10		
Solids Disposal ($.044/kg)		kg/hr			
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	803	kg/hr	418		
Total Direct Operating Cost				627	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			45		
Total Overhead Costs				83	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				192	
TOTAL ANNUALIZED COSTS					902
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	163		
SO_2 Scrubbing	301		
Fans	76		
Wastewater Pumps	25		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	34		
Total Direct Costs (TDC)		599	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	60		
Construction Fees (0.1 TDC)	60		
Start-up (0.02 TDC)	12		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		200	
Contingencies (0.2 (TDC+TIC))		160	
TOTAL TURNKEY COSTS (TTC)			959
Land[b] (0.00084 TTC)		0.8	
Working Capital (0.25 Direct Operating Costs)[c]		148	
TOTAL CAPITAL INVESTMENT (TCI)			1108

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			24		
Maintenance Materials (.04 TDC)			24		
Electricity (25.8 mills/kWh)	139	kW	19		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	24.0	m^3/hr	5		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	17.0	m^3/hr	10		
Solids Disposal ($.044/kg)		kg/hr			
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	741	kg/hr	386		
Total Direct Operating Cost				594	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			45		
Total Overhead Costs				83	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				188	
TOTAL ANNUALIZED COSTS					865
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 118 MW_t (400 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	264		
SO_2 Scrubbing	465		
Fans	126		
Wastewater Pumps	30		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	53		
Total Direct Costs (TDC)		938	
Indirect Capital Costs			
Engineering[a]	94		
Construction and Field Expenses (0.1 TDC)	94		
Construction Fees (0.1 TDC)	94		
Start-up (0.02 TDC)	19		
Performance Test (0.01 TDC)	9		
Total Indirect Cost (TIC)		310	
Contingencies (0.2 (TDC+TIC))		250	
TOTAL TURNKEY COSTS (TTC)			1498
Land[b] (0.00084 TTC)		1.3	
Working Capital (0.25 Direct Operating Costs)[c]		284	
TOTAL CAPITAL INVESTMENT (TCI)			1783

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 118 MW_t (400 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			37		
Maintenance Materials (.04 TDC)			37		
Electricity (25.8 mills/kWh)	298	kW	40		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	54.0	m^3/hr	11		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	42.0	m^3/hr	15		
Solids Disposal ($.044/kg)		kg/hr			
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	1680	kg/hr	873		
Total Direct Operating Cost				1139	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			52		
Total Overhead Costs				90	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				303	
TOTAL ANNUALIZED COSTS					1532
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 2.3% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	84		
SO_2 Scrubbing	183		
Fans	40		
Wastewater Pumps	19		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	20		
Total Direct Costs (TDC)		346	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	35		
Construction Fees (0.1 TDC)	35		
Start-up (0.02 TDC)	7		
Performance Test (0.01 TDC)	3		
Total Indirect Cost (TIC)		142	
Contingencies (0.2 (TDC+TIC))		98	
TOTAL TURNKEY COSTS (TTC)			586
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		71	
TOTAL CAPITAL INVESTMENT (TCI)			657

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 2.3% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			14		
Maintenance Materials (.04 TDC)			14		
Electricity (25.8 mills/kWh)	52.1	kW	7		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	7.26	m^3/hr	2		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	4.54	m^3/hr	6		
Solids Disposal ($.044/kg)		kg/hr			
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	225	kg/hr	117		
Total Direct Operating Cost				286	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			40		
Total Overhead Costs				78	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				112	
TOTAL ANNUALIZED COSTS					476
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 118 MW_t (400 MBtu/hr)
Coal Feedstock: 2.3% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	209		
SO_2 Scrubbing	444		
Fans	127		
Wastewater Pumps	27		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	49		
Total Direct Costs (TDC)		856	
Indirect Capital Costs			
Engineering[a]	94		
Construction and Field Expenses (0.1 TDC)	86		
Construction Fees (0.1 TDC)	86		
Start-up (0.02 TDC)	17		
Performance Test (0.01 TDC)	9		
Total Indirect Cost (TIC)		292	
Contingencies (0.2 (TDC+TIC))		230	
TOTAL TURNKEY COSTS (TTC)			1378
Land[b] (0.00084 TTC)		1.2	
Working Capital (0.25 Direct Operating Costs)[c]		209	
TOTAL CAPITAL INVESTMENT (TCI)			1588

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type:	Sodium Throwaway
Boiler Capacity:	118 MW_t (400 MBtu/hr)
Coal Feedstock:	2.3% S
SO_2 Control Level:	90%
Operating Factor:	60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			34		
Maintenance Materials (.04 TDC)			34		
Electricity (25.8 mills/kWh)	249	kW	34		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	36.8	m^3/hr	8		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	24.5	m^3/hr	12		
Solids Disposal ($.044/kg)		kg/hr			
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	1130	kg/hr	591		
Total Direct Operating Cost				839	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			50		
Total Overhead Costs				88	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				270	
TOTAL ANNUALIZED COSTS					1197
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	37		
SO_2 Scrubbing	107		
Fans	20		
Wastewater Pumps	13		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	11		
Total Direct Costs (TDC)		188	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	19		
Construction Fees (0.1 TDC)	19		
Start-up (0.02 TDC)	4		
Performance Test (0.01 TDC)	2		
Total Indirect Cost (TIC)		106	
Contingencies (0.2 (TDC+TIC))		59	
TOTAL TURNKEY COSTS (TTC)			353
Land[b] (0.00084 TTC)		0.3	
Working Capital (0.25 Direct Operating Costs)[c]		41	
TOTAL CAPITAL INVESTMENT (TCI)			394

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		7		
Maintenance Materials (.04 TDC)		7		
Electricity (25.8 mills/kWh)	17.6 kW	2		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	1.70 m^3/hr	0		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	0.66 m^3/hr	2		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	34.6 kg/hr	18		
Total Direct Operating Cost			162	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		36		
Total Overhead Costs			74	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			67	
TOTAL ANNUALIZED COSTS				303
Annual Unit Costs	\$/10^6 Btu	\$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 85%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	36		
SO_2 Scrubbing	107		
Fans	20		
Wastewater Pumps	13		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	11		
Total Direct Costs (TDC)		187	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	19		
Construction Fees (0.1 TDC)	19		
Start-up (0.02 TDC)	4		
Performance Test (0.01 TDC)	2		
Total Indirect Cost (TIC)		106	
Contingencies (0.2 (TDC+TIC))		59	
TOTAL TURNKEY COSTS (TTC)			352
Land[b] (0.00084 TTC)		0.3	
Working Capital (0.25 Direct Operating Costs)[c]		40	
TOTAL CAPITAL INVESTMENT (TCI)			392

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 85%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		7		
Maintenance Materials (.04 TDC)		7		
Electricity (25.8 mills/kWh)	17.4 kW_e	2		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	1.63 m^3/hr	0		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	0.59 m^3/hr	2		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	33.2 kg/hr	17		
Total Direct Operating Cost			155	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		36		
Total Overhead Costs			74	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			67	
TOTAL ANNUALIZED COSTS				302
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway

Boiler Capacity: 8.8 MW_t (30 MBtu/hr)

Coal Feedstock: 0.6% S Western

SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	36		
SO_2 Scrubbing	107		
Fans	20		
Wastewater Pumps	13		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	11		
Total Direct Costs (TDC)		187	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	19		
Construction Fees (0.1 TDC)	19		
Start-up (0.02 TDC)	4		
Performance Test (0.01 TDC)	2		
Total Indirect Cost (TIC)		106	
Contingencies (0.2 (TDC+TIC))		59	
TOTAL TURNKEY COSTS (TTC)			352
Land[b] (0.00084 TTC)		0.3	
Working Capital (0.25 Direct Operating Costs)[c]		40	
TOTAL CAPITAL INVESTMENT (TCI)			392

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200x10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400x10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 75%
Operating Factor: 60%

Item	Cost (Thousands of dollars)			
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		7		
Maintenance Materials (.04 TDC)		7		
Electricity (25.8 mills/kWh)	17.2 kW	2		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	1.59 m^3/hr	0		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	0.54 m^3/hr	2		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	kg/hr	17		
Total Direct Operating Cost			155	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		36		
Total Overhead Costs			74	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			67	
TOTAL ANNUALIZED COSTS				302
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

---

a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	54		
SO_2 Scrubbing	168		
Fans	39		
Wastewater Pumps	16		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	17		
Total Direct Costs (TDC)		294	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	29		
Construction Fees (0.1 TDC)	29		
Start-up (0.02 TDC)	6		
Performance Test (0.01 TDC)	3		
Total Indirect Cost (TIC)		129	
Contingencies (0.2 (TDC+TIC))		85	
TOTAL TURNKEY COSTS (TTC)			508
Land[b] (0.00084 TTC)		0.4	
Working Capital (0.25 Direct Operating Costs)[c]		51	
TOTAL CAPITAL INVESTMENT (TCI)			559

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Bo.ler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 0.6% S Estern
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		12		
Maintenance Materials (.04 TDC)		12		
Electricity (25.8 mills/kWh)	44.2 kW	6		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	4.20 m^3/hr	1		
Methane ($2.05/GJ)	GJ/hr			
Wastewater Treating[a]	1.66 m^3/hr	4		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	85.4 kg/hr	44		
Total Direct Operating Cost			205	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		39		
Total Overhead Costs			77	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			95	
TOTAL ANNUALIZED COSTS				377
Annual Unit Costs	\$/10^6 Btu	\$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	52		
SO_2 Scrubbing	168		
Fans	39		
Wastewater Pumps	15		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	17		
Total Direct Costs (TDC)		291	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	29		
Construction Fees (0.1 TDC)	29		
Start-up (0.02 TDC)	6		
Performance Test (0.01 TDC)	3		
Total Indirect Cost (TIC)		129	
Contingencies (0.2 (TDC+TIC))		84	
TOTAL TURNKEY COSTS (TTC)			504
Land[b] (0.00084 TTC)		0.4	
Working Capital (0.25 Direct Operating Costs)[c]		50	
TOTAL CAPITAL INVESTMENT (TCI)			554

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 75%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		12		
Maintenance Materials (.04 TDC)		12		
Electricity (25.8 mills/kWh)	43.3 kW	6		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	3.95 m^3/hr	1		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	1.34 m^3/hr	3		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	80.1 kg/hr	42		
Total Direct Operating Cost			202	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		39		
Total Overhead Costs			77	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			95	
TOTAL ANNUALIZED COSTS				374
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

---

a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 44 MW_t (150 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	75		
SO_2 Scrubbing	267		
Fans	66		
Wastewater Pumps	18		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	26		
Total Direct Costs (TDC)		452	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	45		
Construction Fees (0.1 TDC)	45		
Start-up (0.02 TDC)	9		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		166	
Contingencies (0.2 (TDC+TIC))		124	
TOTAL TURNKEY COSTS (TTC)			742
Land[b] (0.00084 TTC)		0.6	
Working Capital (0.25 Direct Operating Costs)[c]		68	
TOTAL CAPITAL INVESTMENT (TCI)			810

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 44 MW_t (150 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		18		
Maintenance Materials (.04 TDC)		18		
Electricity (25.8 mills/kWh)	88.2 kW	12		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	8.51 m^3/hr	2		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	3.31 m^3/hr	5		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	173 kg/hr	90		
Total Direct Operating Cost			271	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		42		
Total Overhead Costs			80	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			138	
TOTAL ANNUALIZED COSTS				489
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 44 MW_t (150 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	72		
SO_2 Scrubbing	267		
Fans	66		
Wastewater Pumps	17		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	25		
Total Direct Costs (TDC)		447	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	45		
Construction Fees (0.1 TDC)	45		
Start-up (0.02 TDC)	9		
Performance Test (0.01 TDC)	4		
Total Indirect Cost (TIC)		165	
Contingencies (0.2 (TDC+TIC))		122	
TOTAL TURNKEY COSTS (TTC)			734
Land[b] (0.00084 TTC)		0.6	
Working Capital (0.25 Direct Operating Costs)[c]		66	
TOTAL CAPITAL INVESTMENT (TCI)			800

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 44 MW_t (150 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 75%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		18		
Maintenance Materials (.04 TDC)		18		
Electricity (25.8 mills/kWh)	86.5 kW	12		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	7.88 m^3/hr	2		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	2.68 m^3/hr	4		
Solids Disposal ($.044/kg)	kg/hr	-		
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	160 kg/hr	83		
Total Direct Operating Cost			263	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		42		
Total Overhead Costs			80	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			136	
TOTAL ANNUALIZED COSTS				479
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	84		
SO_2 Scrubbing	294		
Fans	74		
Wastewater Pumps	19		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	28		
Total Direct Costs (TDC)		499	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	50		
Construction Fees (0.1 TDC)	50		
Start-up (0.02 TDC)	10		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		177	
Contingencies (0.2 (TDC+TIC))		135	
TOTAL TURNKEY COSTS (TTC)			811
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		75	
TOTAL CAPITAL INVESTMENT (TCI)			886

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type:	Sodium Throwaway
Boiler Capacity:	58.6 MW (200 MBtu/hr)
Coal Feedstock:	0.6% S Western
SO_2 Control Level:	90%
Operating Factor:	60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		20		
Maintenance Materials (.04 TDC)		20		
Electricity (25.8 mills/kWh)	104 kW	14		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	10.5 m^3/hr	2		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	4.43 m^3/hr	5		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	214 kg/hr	111		
Total Direct Operating Cost			298	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		43		
Total Overhead Costs			81	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			150	
TOTAL ANNUALIZED COSTS				529
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 85%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	82		
SO_2 Scrubbing	294		
Fans	74		
Wastewater Pumps	19		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	28		
Total Direct Costs (TDC)		497	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	50		
Construction Fees (0.1 TDC)	50		
Start-up (0.02 TDC)	10		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		177	
Contingencies (0.2 (TDC+TIC))		135	
TOTAL TURNKEY COSTS (TTC)			809
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		74	
TOTAL CAPITAL INVESTMENT (TCI)			882

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 85%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		20		
Maintenance Materials (.04 TDC)		20		
Electricity (25.8 mills/kWh)	103 kW	14		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	10.1 m^3/hr	2		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	4.00 m^3/hr	5		
Solids Disposal ($.044/kg)	kg/hr			
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	205 kg/hr	107		
Total Direct Operating Cost			294	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		43		
Total Overhead Costs			81	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			150	
TOTAL ANNUALIZED COSTS				525
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	80		
SO_2 Scrubbing	294		
Fans	74		
Wastewater Pumps	18		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	28		
Total Direct Costs (TDC)		494	
Indirect Capital Costs			
Engineering[a]	62		
Construction and Field Expenses (0.1 TDC)	49		
Construction Fees (0.1 TDC)	49		
Start-up (0.02 TDC)	10		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		175	
Contingencies (0.2 (TDC+TIC))		134	
TOTAL TURNKEY COSTS (TTC)			803
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		72	
TOTAL CAPITAL INVESTMENT (TCI)			875

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a. Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			20		
Maintenance Materials (.04 TDC)			20		
Electricity (25.8 mills/kWh)	102	kW_e	14		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	9.65	m^3/hr	2		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	3.56	m^3/hr	5		
Solids Disposal ($.044/kg)		kg/hr			
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	196	kg/hr	102		
Total Direct Operating Cost				289	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			43		
Total Overhead Costs				81	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				149	
TOTAL ANNUALIZED COSTS					519
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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[a] Reference: 5

[b] Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 118 MW_t (400 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	119		
SO_2 Scrubbing	458		
Fans	124		
Wastewater Pumps	22		
Regeneration	-		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	43		
Total Direct Costs (TDC)		766	
Indirect Capital Costs			
Engineering[a]	94		
Construction and Field Expenses (0.1 TDC)	77		
Construction Fees (0.1 TDC)	77		
Start-up (0.02 TDC)	15		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		271	
Contingencies (0.2 (TDC+TIC))		207	
TOTAL TURNKEY COSTS (TTC)			1244
Land[b] (0.00084 TTC)		1.0	
Working Capital (0.25 Direct Operating Costs)[c]		110	
TOTAL CAPITAL INVESTMENT (TCI)			1355

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Sodium Throwaway
Boiler Capacity: 118 MW_t (400 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			31		
Maintenance Materials (.04 TDC)			31		
Electricity (25.8 mills/kWh)	206	kW	28		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	20.2	m^3/hr	4		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]	8.85	m^3/hr	8		
Solids Disposal ($.044/kg)		kg/hr	---		
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	413	kg/hr	215		
Total Direct Operating Cost				443	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			49		
Total Overhead Costs				87	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				230	
TOTAL ANNUALIZED COSTS					760
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	76		
SO_2 Scrubbing	108		
Fans	20		
Wastewater Pumps	-		
Regeneration	5		
Solids Separation	139		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	21		
Total Direct Costs (TDC)		369	
Indirect Capital Costs			
Engineering[a]	78		
Construction and Field Expenses (0.1 TDC)	37		
Construction Fees (0.1 TDC)	37		
Start-up (0.02 TDC)	7		
Performance Test (0.01 TDC)	4		
Total Indirect Cost (TIC)		163	
Contingencies (0.2 (TDC+TIC))		106	
TOTAL TURNKEY COSTS (TTC)			638
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		60	
TOTAL CAPITAL INVESTMENT (TCI)			699

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		15		
Maintenance Materials (.04 TDC)		15		
Electricity (25.8 mills/kWh)	17.8 kW	2		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	1.30 m^3/hr	0		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	m^3/hr	-		
Solids Disposal ($.044/kg)	283 kg/hr	66		
Chemicals				
Lime ($.0385/kg)	59.5 kg/hr	12		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	5.90 kg/hr	3		
Total Direct Operating Cost			239	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		41		
Total Overhead Costs			79	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			119	
TOTAL ANNUALIZED COSTS				437
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	55		
SO_2 Scrubbing	106		
Fans	20		
Wastewater Pumps	-		
Regeneration	2		
Solids Separation	114		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	18		
Total Direct Costs (TDC)		315	
Indirect Capital Costs			
Engineering[a]	78		
Construction and Field Expenses (0.1 TDC)	32		
Construction Fees (0.1 TDC)	32		
Start-up (0.02 TDC)	6		
Performance Test (0.01 TDC)	3		
Total Indirect Cost (TIC)		151	
Contingencies (0.2 (TDC+TIC))		93	
TOTAL TURNKEY COSTS (TTC)			559
Land[b] (0.00084 TTC)		0.5	
Working Capital (0.25 Direct Operating Costs)[c]		43	
TOTAL CAPITAL INVESTMENT (TCI)			603

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			13		
Maintenance Materials (.04 TDC)			13		
Electricity (25.8 mills/kWh)	16.0	kW_e	2		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	1.03	m^3/hr	0		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]		m^3/hr	-		
Solids Disposal ($.044/kg)	55.4	kg/hr	13		
Chemicals					
Lime ($.0385/kg)	11.4	kg/hr	2		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	2.41	kg/hr	1		
Total Direct Operating Cost				170	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)					
Total Overhead Costs				78	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				102	
TOTAL ANNUALIZED COSTS					350
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FCD Type: Dual Alkali
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	106		
SO_2 Scrubbing	108		
Fans	40		
Wastewater Pumps	-		
Regeneration	10		
Solids Separation	157		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	30		
Total Direct Costs (TDC)		523	
Indirect Capital Costs			
Engineering[a]	78		
Construction and Field Expenses (0.1 TDC)	52		
Construction Fees (0.1 TDC)	52		
Start-up (0.02 TDC)	10		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		197	
Contingencies (0.2 (TDC+TIC))		144	
TOTAL TURNKEY COSTS (TTC)			864
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		95	
TOTAL CAPITAL INVESTMENT (TCI)			960

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		21		
Maintenance Materials (.04 TDC)		21		
Electricity (25.8 mills/kWh)	44.2 kW_e	6		
Steam ($1.84/GJ)	GJ/hr			
Proc. Water ($.04/$m^3$)	3.27 m^3/hr	1		
Methane ($2.05/CJ)	GJ/hr	-		
Wastewater Treating[a]	m^3/hr	-		
Solids Disposal ($.044/kg)	719 kg/hr	166		
Chemicals				
Lime ($.0385/kg)	151 kg/hr	31		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	14.5 kg/hr	8		
Total Direct Operating Cost			380	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		44		
Total Overhead Costs			82	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			163	
TOTAL ANNUALIZED COSTS				625
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 2.3% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	86		
SO_2 Scrubbing	182		
Fans	41		
Wastewater Pumps	-		
Regeneration	7		
Solids Separation	145		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	28		
Total Direct Costs (TDC)		489	
Indirect Capital Costs			
Engineering[a]	78		
Construction and Field Expenses (0.1 TDC)	49		
Construction Fees (0.1 TDC)	49		
Start-up (0.02 TDC)	10		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		191	
Contingencies (0.2 (TDC+TIC))		136	
TOTAL TURNKEY COSTS (TTC)			816
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		72	
TOTAL CAPITAL INVESTMENT (TCI)			889

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a. Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 2.3% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		20		
Maintenance Materials (.04 TDC)		20		
Electricity (25.8 mills/kWh)	42.0 kW_e	6		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	2.87 m^3/hr	1		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	m^3/hr	-		
Solids Disposal ($.044/kg)	403 kg/hr	93		
Chemicals				
Lime ($.0385/kg)	84.4 kg/hr	17		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	8.17 kg/hr	4		
Total Direct Operating Cost			287	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		43		
Total Overhead Costs			81	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			151	
TOTAL ANNUALIZED COSTS				519
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	67		
SO_2 Scrubbing	180		
Fans	41		
Wastewater Pumps	-		
Regeneration	3		
Solids Separation	127		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	25		
Total Direct Costs (TDC)		443	
Indirect Capital Costs			
Engineering[a]	78		
Construction and Field Expenses (0.1 TDC)	44		
Construction Fees (0.1 TDC)	44		
Start-up (0.02 TDC)	9		
Performance Test (0.01 TDC)	4		
Total Indirect Cost (TIC)		179	
Contingencies (0.2 (TDC+TIC))		124	
TOTAL TURNKEY COSTS (TTC)			746
Land[b] (0.00084 TTC)		0.6	
Working Capital (0.25 Direct Operating Costs)[c]		52	
TOTAL CAPITAL INVESTMENT (TCI)			799

---

a. Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		18		
Maintenance Materials (.04 TDC)		18		
Electricity (25.8 mills/kWh)	40.3 kW_e	5		
Steam ($1.84/GJ)	GJ/hr	-		
Proc. Water ($.04/$m^3$)	2.55 m^3/hr	1		
Methane ($2.05/GJ)	GJ/hr	-		
Wastewater Treating[a]	m^3/hr	-		
Solids Disposal ($.044/kg)	137 kg/hr	32		
Chemicals				
Lime ($.0385/kg)	28.6 kg/hr	6		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	5.90 kg/hr	3		
Total Direct Operating Cost			209	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		42		
Total Overhead Costs			80	
By-Product Credits			(-)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			136	
TOTAL ANNUALIZED COSTS				425
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	162		
SO_2 Scrubbing	297		
Fans	76		
Wastewater Pumps	-		
Regeneration	18		
Solids Separation	178		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	44		
Total Direct Costs (TDC)		775	
Indirect Capital Costs			
Engineering[a]	78		
Construction and Field Expenses (0.1 TDC)	78		
Construction Fees (0.1 TDC)	78		
Start-up (0.02 TDC)	16		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		258	
Contingencies (0.2 (TDC+TIC))		207	
TOTAL TURNKEY COSTS (TTC)			1240
Land[b] (0.00084 TTC)		1.0	
Working Capital (0.25 Direct Operating Costs)[c]		181	
TOTAL CAPITAL INVESTMENT (TCI)			1422

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			31		
Maintenance Materials (.04 TDC)			31		
Electricity (25.8 mills/kWh)	105	kW_e	14		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	7.04	m^3/hr	2		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	1820	kg/hr	421		
Chemicals					
Lime ($.0385/kg)	402	kg/hr	81		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	37.2	kg/hr	19		
Total Direct Operating Cost				725	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			49		
Total Overhead Costs				87	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				241	
TOTAL ANNUALIZED COSTS					1053
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	90		
SO_2 Scrubbing	292		
Fans	74		
Wastewater Pumps	-		
Regeneration	6		
Solids Separation	143		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	36		
Total Direct Costs (TDC)		641	
Indirect Capital Costs			
Engineering[a]	78		
Construction and Field Expenses (0.1 TDC)	64		
Construction Fees (0.1 TDC)	64		
Start-up (0.02 TDC)	13		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		225	
Contingencies (0.2 (TDC+TIC))		173	
TOTAL TURNKEY COSTS (TTC)			1039
Land[b] (0.00084 TTC)		0.9	
Working Capital (0.25 Direct Operating Costs)[c]		75	
TOTAL CAPITAL INVESTMENT (TCI)			1115

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a. Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 58.6 MW$_t$ (200 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO$_2$ Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			26		
Maintenance Materials (.04 TDC)			26		
Electricity (25.8 mills/kWh)	93.8	kW$_e$	13		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/m^3)	5.98	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]		m^3/hr	-		
Solids Disposal ($.044/kg)	366	kg/hr	85		
Chemicals					
Lime ($.0385/kg)	76.3	kg/hr	15		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	15.9	kg/hr	8		
Total Direct Operating Cost				300	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			46		
Total Overhead Costs				84	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				190	
TOTAL ANNUALIZED COSTS					574
Annual Unit Costs	$/10^6 Btu		$/kg SO$_2$		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 118 MW (400 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	229		
SO_2 Scrubbing	465		
Fans	126		
Wastewater Pumps	-		
Regeneration	30		
Solids Separation	198		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	63		
Total Direct Costs (TDC)		1110	
Indirect Capital Costs			
Engineering[a]	111		
Construction and Field Expenses (0.1 TDC)	111		
Construction Fees (0.1 TDC)	111		
Start-up (0.02 TDC)	22		
Performance Test (0.01 TDC)	11		
Total Indirect Cost (TIC)		366	
Contingencies (0.2 (TDC+TIC))		295	
TOTAL TURNKEY COSTS (TTC)			1771
Land[b] (0.00084 TTC)		1.5	
Working Capital (0.25 Direct Operating Costs)[c]		334	
TOTAL CAPITAL INVESTMENT (TCI)			2105

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali

Boiler Capacity: 118 MW_t (400 MBtu/hr)

Coal Feedstock: 3.5% S Eastern

SO_2 Control Level: 90%

Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			43		
Maintenance Materials (.04 TDC)			43		
Electricity (25.8 mills/kWh)	207	kW_e	28		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	14.0	m^3/hr	3		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]		m^3/hr	-		
Solids Disposal ($.044/kg)	3840	kg/hr	887		
Chemicals					
Lime ($.0385/kg)	806	kg/hr	163		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	78.5	kg/hr	41		
Total Direct Operating Cost				1334	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			55		
Total Overhead Costs				93	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				351	
TOTAL ANNUALIZED COSTS					1778
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 118 MW$_t$ (400 MBtu/hr)
Coal Feedstock: 2.3% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	173		
SO_2 Scrubbing	440		
Fans	127		
Wastewater Pumps	-		
Regeneration	20		
Solids Separation	183		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	57		
Total Direct Costs (TDC)		1000	
Indirect Capital Costs			
Engineering[a]	108		
Construction and Field Expenses (0.1 TDC)	100		
Construction Fees (0.1 TDC)	100		
Start-up (0.02 TDC)	20		
Performance Test (0.01 TDC)	10		
Total Indirect Cost (TIC)		338	
Contingencies (0.2 (TDC+TIC))		268	
TOTAL TURNKEY COSTS (TTC)			1606
Land[b] (0.00084 TTC)		1.3	
Working Capital (0.25 Direct Operating Costs)[c]		212	
TOTAL CAPITAL INVESTMENT (TCI)			1819

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 118 MW_t (400 MBtu/hr)
Coal Feedstock: 2.3% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			40		
Maintenance Materials (.04 TDC)			40		
Electricity (25.8 mills/kWh)	195	kW_e	26		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/m^3)	13.3	m^3/hr	3		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]		m^3/hr	-		
Solids Disposal ($.044/kg)	2150	kg/hr	497		
Chemicals					
Lime ($.0385/kg)	449	kg/hr	91		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	44.9	kg/hr	23		
Total Direct Operating Cost				846	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			54		
Total Overhead Costs				92	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				309	
TOTAL ANNUALIZED COSTS					1247
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 118 MW_t (400 MBtu/hr
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	118		
SO_2 Scrubbing	453		
Fans	124		
Wastewater Pumps	-		
Regeneration	10		
Solids Separation	155		
Solids Collection	-		
Purge Treatment	-		
Sulfur Production	-		
Utilities and Services	52		
Total Direct Costs (TDC)		912	
Indirect Capital Costs			
Engineering[a]	108		
Construction and Field Expenses (0.1 TDC)	91		
Construction Fees (0.1 TDC)	91		
Start-up (0.02 TDC)	18		
Performance Test (0.01 TDC)	9		
Total Indirect Cost (TIC)		317	
Contingencies (0.2 (TDC+TIC))		246	
TOTAL TURNKEY COSTS (TTC)			1475
Land[b] (0.00084 TTC)		1.2	
Working Capital (0.25 Direct Operating Costs)[c]		111	
TOTAL CAPITAL INVESTMENT (TCI)			1587

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Dual Alkali
Boiler Capacity: 118 MW_t (400 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			36		
Maintenance Materials (.04 TDC)			36		
Electricity (25.8 mills/kWh)	185	kW_e	25		
Steam ($1.84/GJ)		GJ/hr	-		
Proc. Water ($.04/$m^3$)	11.8	m^3/hr	2		
Methane ($2.05/GJ)		GJ/hr	-		
Wastewater Treating[a]		m^3/hr	-		
Solids Disposal ($.044/kg)	733	kg/hr	169		
Chemicals					
Lime ($.0385/kg)	153	kg/hr	31		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	32.2	kg/hr	17		
Total Direct Operating Cost				442	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			51		
Total Overhead Costs				89	
By-Product Credits				(-)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				269	
TOTAL ANNUALIZED COSTS					800
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	79		
SO_2 Scrubbing	230		
Fans	66		
Wastewater Pumps			
Regeneration			
Solids Separation	426		
Solids Collection			
Purge Treatment			
Sulfur Production			
Utilities and Services	48		
Total Direct Costs (TDC)		850	
Indirect Capital Costs			
Engineering[a]	92		
Construction and Field Expenses (0.1 TDC)	85		
Construction Fees (0.1 TDC)	85		
Start-up (0.02 TDC)	17		
Performance Test (0.01 TDC)	9		
Total Indirect Cost (TIC)		288	
Contingencies (0.2 (TDC+TIC))		228	
TOTAL TURNKEY COSTS (TTC)			1366
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		64	
TOTAL CAPITAL INVESTMENT (TCI)			1431

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			34		
Maintenance Materials (.04 TDC)			34		
Electricity (25.8 mills/kWh)	92	kW_e	12		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	4.4	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	163	kg/hr	38		
Chemicals					
Lime ($.0385/kg)	85	kg/hr	17		
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				261	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			50		
Total Overhead Costs				88	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				242	
TOTAL ANNUALIZED COSTS					592
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	65		
SO_2 Scrubbing	230		
Fans	66		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	426		
Purge Treatment			
Sulfur Production			
Utilities and Services	47		
Total Direct Costs (TDC)		834	
Indirect Capital Costs			
Engineering[a]	92		
Construction and Field Expenses (0.1 TDC)	83		
Construction Fees (0.1 TDC)	83		
Start-up (0.02 TDC)	17		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		283	
Contingencies (0.2 (TDC+TIC))		223	
TOTAL TURNKEY COSTS (TTC)			1340
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		79	
TOTAL CAPITAL INVESTMENT (TCI)			1420

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			33		
Maintenance Materials (.04 TDC)			33		
Electricity (25.8 mills/kWh)	92	kW_e	13		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	4.4	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	155	kg/hr	36		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)	141	kg/hr	74		
Total Direct Operating Cost				316	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			50		
Total Overhead Costs				88	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				240	
TOTAL ANNUALIZED COSTS					644
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: 200×10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	75		
SO_2 Scrubbing	247		
Fans	73		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	470		
Purge Treatment			
Sulfur Production			
Utilities and Services	59		
Total Direct Costs (TDC)		917	
Indirect Capital Costs			
Engineering[a]	92		
Construction and Field Expenses (0.1 TDC)	92		
Construction Fees (0.1 TDC)	92		
Start-up (0.02 TDC)	18		
Performance Test (0.01 TDC)	9		
Total Indirect Cost (TIC)		303	
Contingencies (0.2 (TDC+TIC))		244	
TOTAL TURNKEY COSTS (TTC)			1464
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		91	
TOTAL CAPITAL INVESTMENT (TCI)			1556

[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: 200×10^6 Btu/hr (58.6 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			37		
Maintenance Materials (.04 TDC)			37		
Electricity (25.8 mills/kWh)	105	kW_e	14		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	5.2	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	219	kg/hr	51		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)	188	kg/hr	98		
Total Direct Operating Cost				364	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			52		
Total Overhead Costs				90	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				264	
TOTAL ANNUALIZED COSTS					718
Annual Unit Costs	\$/10^6 Btu		\$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)
Boiler Capacity: 400×10^6 Btu/hr (117 MW_t)
Coal Feedstock: Eastern 2.3% S
SO_2 Control Level: 70%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	179		
SO_2 Scrubbing	349		
Fans	126		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	830		
Purge Treatment			
Sulfur Production			
Utilities and Services	89		
Total Direct Costs (TDC)		1,573	
Indirect Capital Costs			
Engineering[a]	157		
Construction and Field Expenses (0.1 TDC)	157		
Construction Fees (0.1 TDC)	157		
Start-up (0.02 TDC)	31		
Performance Test (0.01 TDC)	16		
Total Indirect Cost (TIC)		518	
Contingencies (0.2 (TDC+TIC))		418	
TOTAL TURNKEY COSTS (TTC)			2509
Land[b] (0.00084 TTC)		2	
Working Capital (0.25 Direct Operating Costs)[c]		171	
TOTAL CAPITAL INVESTMENT (TCI)			2682

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)
Boiler Capacity: 400×10^6 Btu/hr (118 MW_t)
Coal Feedstock: Eastern 2.3% S
SO_2 Control Level: 70%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			63		
Maintenance Materials (.04 TDC)			63		
Electricity (25.8 mills/kWh)	209	kW_e	28		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	10	m^3/hr	2		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	1181	kg/hr	273		
Chemicals					
Lime ($.0385/kg)	631	kg/hr	128		
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				683	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			66		
Total Overhead Costs				104	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				455	
TOTAL ANNUALIZED COSTS					1242
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	75		
SO_2 Scrubbing	230		
Fans	66		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	426		
Purge Treatment			
Sulfur Production			
Utilities and Services	48		
Total Direct Costs (TDC)		845	
Indirect Capital Costs			
Engineering[a]	92		
Construction and Field Expenses (0.1 TDC)	85		
Construction Fees (0.1 TDC)	85		
Start-up (0.02 TDC)	17		
Performance Test (0.01 TDC)	9		
Total Indirect Cost (TIC)		288	
Contingencies (0.2 (TDC+TIC))		226	
TOTAL TURNKEY COSTS (TTC)			1359
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		90	
TOTAL CAPITAL INVESTMENT (TCI)			1450

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			34		
Maintenance Materials (.04 TDC)			34		
Electricity (25.8 mills/kWh)	93	kW_e	13		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	4.4	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	227	kg/hr	53		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)	194	kg/hr	101		
Total Direct Operating Cost				362	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			50		
Total Overhead Costs				88	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				245	
TOTAL ANNUALIZED COSTS					695
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 50%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	56		
SO_2 Scrubbing	230		
Fans	66		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	426		
Purge Treatment			
Sulfur Production			
Utilities and Services	47		
Total Direct Costs (TDC)		826	
Indirect Capital Costs			
Engineering[a]	92		
Construction and Field Expenses (0.1 TDC)	83		
Construction Fees (0.1 TDC)	83		
Start-up (0.02 TDC)	16		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		282	
Contingencies (0.2 (TDC+TIC))		222	
TOTAL TURNKEY COSTS (TTC)			1330
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		70	
TOTAL CAPITAL INVESTMENT (TCI)			1401

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 50%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			33		
Maintenance Materials (.04 TDC)			33		
Electricity (25.8 mills/kWh)	92	kW_e	13		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	4.4	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	109	kg/hr	25		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)	97	kg/hr	51		
Total Direct Operating Cost				282	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			50		
Total Overhead Costs				88	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				237	
TOTAL ANNUALIZED COSTS					607
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type. Spray Drying (Lime)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	94		
SO_2 Scrubbing	230		
Fans	66		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	426		
Purge Treatment			
Sulfur Production			
Utilities and Services	49		
Total Direct Costs (TDC)		865	
Indirect Capital Costs			
Engineering[a]	92		
Construction and Field Expenses (0.1 TDC)	86		
Construction Fees (0.1 TDC)	86		
Start-up (0.02 TDC)	17		
Performance Test (0.01 TDC)	9		
Total Indirect Cost (TIC)		290	
Contingencies (0.2 (TDC+TIC))		231	
TOTAL TURNKEY COSTS (TTC)			1386
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		73	
TOTAL CAPITAL INVESTMENT (TCI)			1460

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			35		
Maintenance Materials (.04 TDC)			35		
Electricity (25.8 mills/kWh)	93	kW_e	13		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	4.4	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	236	kg/hr	55		
Chemicals					
Lime ($.0385/kg)	141	kg/hr	29		
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				294	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			51		
Total Overhead Costs				89	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				247	
TOTAL ANNUALIZED COSTS					630
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)
Boiler Capacity: 150×10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 50%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	69		
SO_2 Scrubbing	230		
Fans	66		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	426		
Purge Treatment			
Sulfur Production			
Utilities and Services	47		
Total Direct Costs (TDC)		839	
Indirect Capital Costs			
Engineering[a]	92		
Construction and Field Expenses (0.1 TDC)	84		
Construction Fees (0.1 TDC)	84		
Start-up (0.02 TDC)	17		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		285	
Contingencies (0.2 (TDC+TIC))		225	
TOTAL TURNKEY COSTS (TTC)			1349
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		58	
TOTAL CAPITAL INVESTMENT (TCI)			1408

a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)
Boiler Capacity: 150x10^6 Btu/hr (44 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 50%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			34		
Maintenance Materials (.04 TDC)			34		
Electricity (25.8 mills/kWh)	92	kW_e	12		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	4.4	m^3/hr	1		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	99	kg/hr	23		
Chemicals					
Lime ($.0385/kg)	46	kg/hr	9		
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				239	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			50		
Total Overhead Costs				88	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				238	
TOTAL ANNUALIZED COSTS					565
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)

Boiler Capacity: 400×10^6 Btu/hr (118 MW_t)

Coal Feedstock: Western 0.6% S

SO_2 Control Level: 70%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	116		
SO_2 Scrubbing	337		
Fans	124		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	840		
Purge Treatment			
Sulfur Production			
Utilities and Services	85		
Total Direct Costs (TDC)		1501	
Indirect Capital Costs			
Engineering[a]	150		
Construction and Field Expenses (0.1 TDC)	150		
Construction Fees (0.1 TDC)	150		
Start-up (0.02 TDC)	30		
Performance Test (0.01 TDC)	15		
Total Indirect Cost (TIC)		495	
Contingencies (0.2 (TDC+TIC))		399	
TOTAL TURNKEY COSTS (TTC)			2395
Land[b] (0.00084 TTC)		2	
Working Capital (0.25 Direct Operating Costs)[c]		106	
TOTAL CAPITAL INVESTMENT (TCI)			2503

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)
Boiler Capacity: 400×10^6 Btu/hr (118 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 70%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			60		
Maintenance Materials (.04 TDC)			60		
Electricity (25.8 mills/kWh)	203	kW_e	27		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	10	m^3/hr	2		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	424	kg/hr	98		
Chemicals					
Lime ($.0385/kg)	227	kg/hr	46		
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				419	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			64		
Total Overhead Costs				102	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				425	
TOTAL ANNUALIZED COSTS					946
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: 60×10^6 Btu/hr (17 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	42		
SO_2 Scrubbing	149		
Fans	33		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	200		
Purge Treatment			
Sulfur Production			
Utilities and Services	25		
Total Direct Costs (TDC)		450	
Indirect Capital Costs			
Engineering[a]	92		
Construction and Field Expenses (0.1 TDC)	45		
Construction Fees (0.1 TDC)	45		
Start-up (0.02 TDC)	9		
Performance Test (0.01 TDC)	5		
Total Indirect Cost (TIC)		196	
Contingencies (0.2 (TDC+TIC))		129	
TOTAL TURNKEY COSTS (TTC)			775
Land[b] (0.00084 TTC)		0.6	
Working Capital (0.25 Direct Operating Costs)[c]		53	
TOTAL CAPITAL INVESTMENT (TCI)			828

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Sodium)
Boiler Capacity: $60x10^6$ Btu/hr (17 MW_t)
Coal Feedstock: Western 0.6% S
SO_2 Control Level: 75%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			18		
Maintenance Materials (.04 TDC)			18		
Electricity (25.8 mills/kWh)	45	kW_e	6		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water (\$.04/m^3)	1.8	m^3/hr	0.4		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	65	kg/hr	15		
Chemicals					
Lime ($.0385/kg)		kg/hr			
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 (\$.0991/kg)	56	kg/hr	29		
Total Direct Operating Cost				212	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			42		
Total Overhead Costs				80	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				140	
TOTAL ANNUALIZED COSTS					432
Annual Unit Costs	\$/10^6 Btu		\$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Spray Drying
Boiler Capacity: 75×10^6 Btu/hr (22 MW_t)
Coal Feedstock: Eastern 2.3% S
SO_2 Control Level: 70%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	87		
SO_2 Scrubbing	171		
Fans	41		
Wastewater Pumps			
Regeneration			
Solids Separation			
Solids Collection	250		
Purge Treatment			
Sulfur Production			
Utilities and Services	32		
Total Direct Costs (TDC)		582	
Indirect Capital Costs			
Engineering[a]	92		
Construction and Field Expenses (0.1 TDC)	58		
Construction Fees (0.1 TDC)	58		
Start-up (0.02 TDC)	12		
Performance Test (0.01 TDC)	6		
Total Indirect Cost (TIC)		226	
Contingencies (0.2 (TDC+TIC))		162	
TOTAL TURNKEY COSTS (TTC)			970
Land[b] (0.00084 TTC)		1	
Working Capital (0.25 Direct Operating Costs)[c]		65	
TOTAL CAPITAL INVESTMENT (TCI)			1036

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Spray Drying (Lime)
Boiler Capacity: 75×10^6 Btu/hr (22 MW_t)
Coal Feedstock: Eastern 2.3% S
SO_2 Control Level: 70%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			25		
Maintenance Materials (.04 TDC)			25		
Electricity (25.8 mills/kWh)	62	kW_e	8		
Steam ($1.84/GJ)		GJ/hr			
Proc. Water ($.04/$m^3$)	2.2	m^3/hr	0.5		
Methane ($2.05/GJ)		GJ/hr			
Wastewater Treating[a]		m^3/hr			
Solids Disposal ($.044/kg)	222	kg/hr	52		
Chemicals					
Lime ($.0385/kg)	119	kg/hr	24		
Limestone ($.0143/kg)		kg/hr			
Na_2CO_3 ($.0991/kg)		kg/hr			
Total Direct Operating Cost				260	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			46		
Total Overhead Costs				84	
By-Product Credits				(0)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				176	
TOTAL ANNUALIZED COSTS					520
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	23		
SO_2 Scrubbing	104		
Fans	32		
Wastewater Pumps	-		
Regeneration	176		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	115		
Sulfur Production	298		
Utilities and Services	48		
Total Direct Costs (TDC)		796	
Indirect Capital Costs			
Engineering[a]	257		
Construction and Field Expenses (0.1 TDC)	80		
Construction Fees (0.1 TDC)	80		
Start-up (0.02 TDC)	16		
Performance Test (0.01 TDC)	8		
Total Indirect Cost (TIC)		441	
Contingencies (0.2 (TDC+TIC))		247	
TOTAL TURNKEY COSTS (TTC)			1484
Land[b] (0.00084 TTC)		1.2	
Working Capital (0.25 Direct Operating Costs)[c]		54	
TOTAL CAPITAL INVESTMENT (TCI)			1539

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			32		
Maintenance Materials (.04 TDC)			32		
Electricity (25.8 mills/kWh)	17	kW_e	2		
Steam ($1.84/GJ)	1.2	GJ/hr	12		
Proc. Water ($.04/$m^3$)	4.1	m^3/hr	1		
Methane ($2.05/GJ)	0.6	GJ/hr	6		
Wastewater Treating[a]		m^3/hr	-		
Solids Disposal ($.044/kg)		kg/hr	-		
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	12	kg/hr	6		
Total Direct Operating Cost				217	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			49		
Total Overhead Costs				87	
By-Product Credits				(-8)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				262	
TOTAL ANNUALIZED COSTS					558
Annual Unit Costs	$/$10^6$ Btu		$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	9		
SO_2 Scrubbing	104		
Fans	32		
Wastewater Pumps	-		
Regeneration	67		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	46		
Sulfur Production	128		
Utilities and Services	16		
Total Direct Costs (TDC)		370	
Indirect Capital Costs			
Engineering[a]	257		
Construction and Field Expenses (0.1 TDC)	37		
Construction Fees (0.1 TDC)	37		
Start-up (0.02 TDC)	7		
Performance Test (0.01 TDC)	4		
Total Indirect Cost (TIC)		342	
Contingencies (0.2 (TDC+TIC))		142	
TOTAL TURNKEY COSTS (TTC)			854
Land[b] (0.00084 TTC)		0.7	
Working Capital (0.25 Direct Operating Costs)[c]		41	
TOTAL CAPITAL INVESTMENT (TCI)			896

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 8.8 MW_t (30 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			15		
Maintenance Materials (.04 TDC)			15		
Electricity (25.8 mills/kWh)	15	kW_e	2		
Steam (\$1.84/GJ)	0.25	GJ/hr	2		
Proc. Water (\$.04/m^3)	1.7	m^3/hr	0		
Methane (\$2.05/GJ)	0.13	GJ/hr	1		
Wastewater Treating[a]		m^3/hr	-		
Solids Disposal (\$.044/kg)		kg/hr	-		
Chemicals					
Lime (\$.0385/kg)		kg/hr	-		
Limestone (\$.0143/kg)		kg/hr	-		
Na_2CO_3 (\$.0991/kg)	2	kg/hr	1		
Total Direct Operating Cost				162	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			41		
Total Overhead Costs				79	
By-Product Credits				(-8)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				152	
TOTAL ANNUALIZED COSTS					385
Annual Unit Costs	$\$/10^6$ Btu		$\$/kg\ SO_2$		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	38		
SO_2 Scrubbing	217		
Fans	59		
Wastewater Pumps	-		
Regeneration	317		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	199		
Sulfur Production	497		
Utilities and Services	93		
Total Direct Costs (TDC)		1420	
Indirect Capital Costs			
Engineering[a]	257		
Construction and Field Expenses (0.1 TDC)	142		
Construction Fees (0.1 TDC)	142		
Start-up (0.02 TDC)	28		
Performance Test (0.01 TDC)	14		
Total Indirect Cost (TIC)		583	
Contingencies (0.2 (TDC+TIC))		401	
TOTAL TURNKEY COSTS (TTC)			2404
Land[b] (0.00084 TTC)		2.0	
Working Capital (0.25 Direct Operating Costs)[c]		77	
TOTAL CAPITAL INVESTMENT (TCI)			2483

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[a] Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

[b] Reference: 4

[c] From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 22 MW_t (75 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		57		
Maintenance Materials (.04 TDC)		57		
Electricity (25.8 mills/kWh)	43.5 kW_e	6		
Steam ($1.84/GJ)	3.0 GJ/hr	29		
Proc. Water ($.04/$m^3$)	4.5 m^3/hr	1		
Methane ($2.05/GJ)	1.4 GJ/hr	15		
Wastewater Treating[a]	m^3/hr	-		
Solids Disposal ($.044/kg)	kg/hr	-		
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	30 kg/hr	16		
Total Direct Operating Cost			307	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		62		
Total Overhead Costs			100	
By-Product Credits			(-20)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			422	
TOTAL ANNUALIZED COSTS				809
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 22 MW (75 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	16		
SO_2 Scrubbing	217		
Fans	59		
Wastewater Pumps	-		
Regeneration	119		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	79		
Sulfur Production	211		
Utilities and Services	31		
Total Direct Costs (TDC)		732	
Indirect Capital Costs			
Engineering[a]	257		
Construction and Field Expenses (0.1 TDC)	73		
Construction Fees (0.1 TDC)	73		
Start-up (0.02 TDC)	15		
Performance Test (0.01 TDC)	7		
Total Indirect Cost (TIC)		425	
Contingencies (0.2 (TDC+TIC))		231	
TOTAL TURNKEY COSTS (TTC)			1388
Land[b] (0.00084 TTC)		1.2	
Working Capital (0.25 Direct Operating Costs)[c]		51	
TOTAL CAPITAL INVESTMENT (TCI)			1440

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200×10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400×10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 22 MW$_t$ (75 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		29		
Maintenance Materials (.04 TDC)		29		
Electricity (25.8 mills/kWh)	38.1 kW$_e$	5		
Steam ($1.84/GJ)	0.64 GJ/hr	6		
Proc. Water ($.04/$m^3$)	3.4 m^3/hr	1		
Methane ($2.05/GJ)	0.30 GJ/hr	3		
Wastewater Treating[a]	m^3/hr	-		
Solids Disposal ($.044/kg)	kg/hr	-		
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	6 kg/hr	3		
Total Direct Operating Cost			202	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		48		
Total Overhead Costs			86	
By-Product Credits			(-4)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			245	
TOTAL ANNUALIZED COSTS				529
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 58.6 MW$_t$ (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	58		
SO_2 Scrubbing	432		
Fans	106		
Wastewater Pumps	-		
Regeneration	587		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	354		
Sulfur Production	852		
Utilities and Services	184		
Total Direct Costs (TDC)		2573	
Indirect Capital Costs			
Engineering[a]	257		
Construction and Field Expenses (0.1 TDC)	257		
Construction Fees (0.1 TDC)	257		
Start-up (0.02 TDC)	51		
Performance Test (0.01 TDC)	26		
Total Indirect Cost (TIC)		848	
Contingencies (0.2 (TDC+TIC))		684	
TOTAL TURNKEY COSTS (TTC)			4105
Land[b] (0.00084 TTC)		3.4	
Working Capital (0.25 Direct Operating Costs)[c]		125	
TOTAL CAPITAL INVESTMENT (TCI)			4233

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a. Engineering Costs = 0.1 TDC for the 58.6 MW (200x10^6 Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW (400x10^6 Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 3.5% S Eastern
SO_2 Control Level: 90%
Operating Factor: 60%

Item			Cost (Thousands of dollars)		
Direct Costs					
Operating Labor (12.02/m-h)			105		
Supervision (15.63/m-h)			21		
Maintenance Labor (.04 TDC)			103		
Maintenance Materials (.04 TDC)			103		
Electricity (25.8 mills/kWh)	105	kW_e	14		
Steam ($1.84/GJ)	8.0	GJ/hr	77		
Proc. Water ($\$.04/m^3$)	9.6	m^3/hr	2		
Methane ($2.05/GJ)	3.8	GJ/hr	41		
Wastewater Treating[a]		m^3/hr	-		
Solids Disposal ($.044/kg)		kg/hr	-		
Chemicals					
Lime ($.0385/kg)		kg/hr	-		
Limestone ($.0143/kg)		kg/hr	-		
Na_2CO_3 ($.0991/kg)	61	kg/hr	32		
Total Direct Operating Cost				498	
Overhead					
Payroll (.3x(1+2) above)			38		
Plant (.26x(1+2+3+4) above)			86		
Total Overhead Costs				124	
By-Product Credits				(-53)	
Capital Charges[b]					
Capital Recovery (.17 TCI)				720	
TOTAL ANNUALIZED COSTS					1289
Annual Unit Costs	$\$/10^6$ Btu		$\$/kg\ SO_2$		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.

CAPITAL INVESTMENT COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%

Item	Cost (Thousands of dollars)		
Direct Capital Costs			
Raw Material Handling	24		
SO_2 Scrubbing	438		
Fans	107		
Wastewater Pumps	-		
Regeneration	220		
Solids Separation	-		
Solids Collection	-		
Purge Treatment	141		
Sulfur Production	362		
Utilities and Services	62		
Total Direct Costs (TDC)		1354	
Indirect Capital Costs			
Engineering[a]	257		
Construction and Field Expenses (0.1 TDC)	135		
Construction Fees (0.1 TDC)	135		
Start-up (0.02 TDC)	27		
Performance Test (0.01 TDC)	14		
Total Indirect Cost (TIC)		568	
Contingencies (0.2 (TDC+TIC))		384	
TOTAL TURNKEY COSTS (TTC)			2306
Land[b] (0.00084 TTC)		1.9	
Working Capital (0.25 Direct Operating Costs)[c]		70	
TOTAL CAPITAL INVESTMENT (TCI)			2378

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a. Engineering Costs = 0.1 TDC for the 58.6 MW ($200x10^6$ Btu/hr) case burning 3.5% S coal @ 90% SO_2 removal. This cost remains constant for the smaller cases. For the 118 MW ($400x10^6$ Btu/hr) cases, the engineering cost is constant and equal to that case burning 3.5% S coal @ 90% SO_2 removal.

b. Reference: 4

c. From Annual Cost Table

ANNUALIZED COSTS FOR FGD PROCESSES

FGD Type: Wellman-Lord
Boiler Capacity: 58.6 MW_t (200 MBtu/hr)
Coal Feedstock: 0.6% S Western
SO_2 Control Level: 90%
Operating Factor: 60%

Item		Cost (Thousands of dollars)		
Direct Costs				
Operating Labor (12.02/m-h)		105		
Supervision (15.63/m-h)		21		
Maintenance Labor (.04 TDC)		54		
Maintenance Materials (.04 TDC)		54		
Electricity (25.8 mills/kWh)	89.7 kW_e	12		
Steam ($1.84/GJ)	1.7 GJ/hr	16		
Proc. Water ($.04/$m^3$)	7.2 m^3/hr	2		
Methane ($2.05/GJ)	0.79 GJ/hr	8		
Wastewater Treating[a]	m^3/hr	-		
Solids Disposal ($.044/kg)	kg/hr	-		
Chemicals				
Lime ($.0385/kg)	kg/hr	-		
Limestone ($.0143/kg)	kg/hr	-		
Na_2CO_3 ($.0991/kg)	13 kg/hr	7		
Total Direct Operating Cost			279	
Overhead				
Payroll (.3x(1+2) above)		38		
Plant (.26x(1+2+3+4) above)		61		
Total Overhead Costs			99	
By-Product Credits			(-11)	
Capital Charges[b]				
Capital Recovery (.17 TCI)			404	
TOTAL ANNUALIZED COSTS				771
Annual Unit Costs	$/$10^6$ Btu	$/kg SO_2		

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a. Reference: 5

b. Capital charges include investment return of 10%; straight-line depreciation, (15 yrs. FGD life) G&A, property taxes and insurance. Federal and State income taxes are not included.